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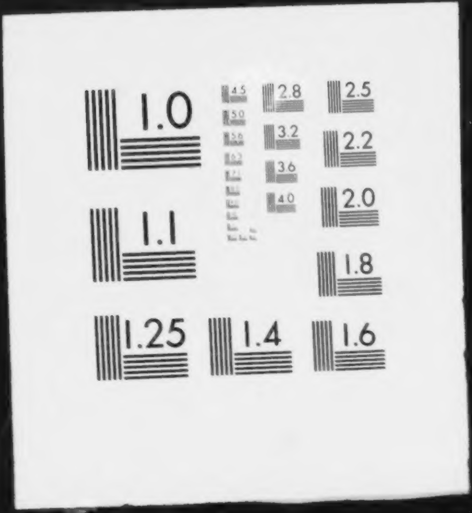

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Lurgi Oil Shale Retorting with Open Pit Mining

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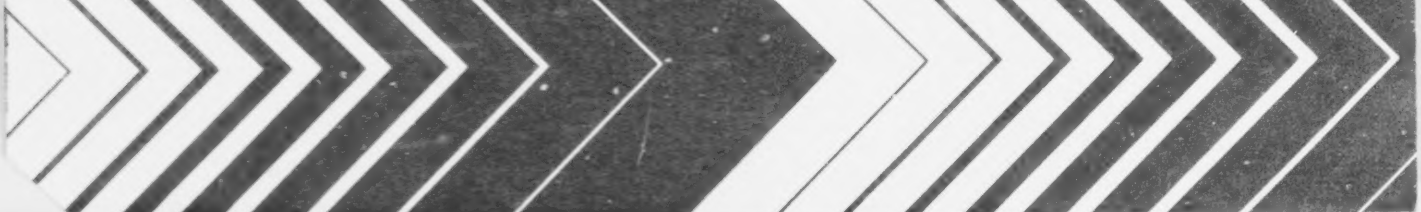

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The Lurgi oil shale PCTM addresses the Lurgi retorting technology, developed by Lurgi Kohle and Mineralotechnik GmbH, West Germany, in the manner in which this technology may be applied to the oil shales of the western United States. This process has been proposed for use by both the Rio Blanco Oil Shale Company (a partnership of Gulf Oil Corporation and Standard Oil Company [Indiana]) and Occidental Oil Shale, Inc., in the phased development of their Federal oil shale lease Tracts C-a and C-b in western Colorado. This document describes a commercial-scale Lurgi oil shale plant, coupled with an open pit mine, based on the design proposed by Rio Blanco Oil Shale Company.

This manual proceeds through a description of the Lurgi oil shale plant proposed by Rio Blanco Oil Shale Company, characterizes the waste streams produced in each medium, and discusses the array of commercially available controls which can be applied to the Lurgi plant waste streams. From these generally characterized controls, several are examined in more detail for each medium in order to illustrate typical control technology operation. Control technology cost and performance estimates are presented, together with descriptions of the discharge streams, secondary waste streams and energy requirements. A summary of data limitations and needs for environmental and control technology considerations is presented.

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Program Manager: Gregory G. Ondich
Office of Environmental Engineering and Technology (RD-681)
U.S. Environmental Protection Agency
401 M Street, SW
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Project Officer: Edward R. Bates
Industrial Environmental Research Laboratory - Cincinnati
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The information in this document has been funded wholly or in part by the United States Environmental Protection Agency under Cooperative Agreement CR-807294 to the Denver Research Institute, University of Denver. It has been subject to the Agency's peer and administrative review, and it has been approved for publication as an EPA document. Mention of trade names or commercial products does not constitute endorsement or recommendation for use.

FOREWORD

The purpose of the Pollution Control Technical Manuals (PCTMs) is to provide process, discharge, and pollution control data in summarized form for the use of permit writers, developers, and other interested parties. The PCTM series covers a range of alternate fuel sources, including coal gasification, coal liquefaction by direct and indirect processing, and the retorting of oil shale.

The series consists of a set of technical volumes directed at production facilities based upon specific conversion processes. The entire series is supplemented by an appendix volume which describes the operation and application of approximately 50 control processes.

All PCTMs are prepared on a base plant concept (coal gasification and liquefaction) or developers' proposed designs (oil shale) which may not fully reflect plants to be built in the future. The PCTMs present examples of control applications, both as individual process units and as integrated control trains. These examples are taken in part from applicable permit applications and, therefore, are reflective of specific plants. None of the examples are intended to convey an Agency endorsement or recommendation, but rather are presented for illustrative purposes. The selection of control technologies for application to specific plants is the exclusive function of the designers and permitters who have the flexibility to utilize the lowest cost and/or most effective approaches. It is hoped that readers will be able to relate their waste streams and controls to those presented in these manuals to enable them to better understand the extent to which various technologies may control specific waste streams and utilize the information in making control technology selections for their specific needs.

The reader should be aware that the PCTMs contain no legally binding requirements or guidance, and that nothing contained in the PCTMs relieves a facility from compliance with existing or future environmental regulations or permit requirements.

H. Wiser

Herbert L. Wiser
Acting Deputy Assistant Administrator
Office of Research and Development
U.S. Environmental Protection Agency

ABSTRACT

The Environmental Protection Agency (EPA), Office of Research and Development, has undertaken an extensive study to determine synthetic fuel plant waste stream characteristics and pollution control systems. The purpose of this and all other PCTMs is to convey this information in a manner that is readily useful to designers, permit writers and the public.

The Lurgi oil shale PCTM addresses the Lurgi retorting technology, developed by Lurgi Kohle und Mineralötechnik GmbH, West Germany, in the manner in which this technology may be applied to the oil shales of the western United States. This process has been proposed for use by both the Rio Blanco Oil Shale Company (a partnership of Gulf Oil Corporation and Standard Oil Company [Indiana]) and Cathedral Bluffs Shale Oil Company (a partnership of Occidental Oil Shale, Inc. and Tenneco Shale Oil Company) in the phased development of their Federal oil shale lease Tracts C-a and C-b in western Colorado. This document describes a commercial-scale Lurgi oil shale plant, coupled with an open pit mine, based on the design proposed by Rio Blanco Oil Shale Company. Plants proposed or built by other developers in the future can be expected to be similar in most aspects to the plant described in this document, but each can be expected to vary in some respects, such as mining methods, selection of particular control technologies, or methods for upgrading the raw shale oil.

This manual proceeds through a description of the Lurgi oil shale plant proposed by Rio Blanco Oil Shale Company, characterizes the waste streams produced in each medium, and discusses the array of commercially available controls which can be applied to the Lurgi plant waste streams. From these generally characterized controls, several are examined in more detail for each medium in order to illustrate typical control technology operation. Control technology cost and performance estimates are presented, together with descriptions of the discharge streams, secondary waste streams and energy requirements. A summary of data limitations and needs for environmental and control technology considerations is presented.

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ABBREVIATIONS

°API	-- gravity (American Petroleum Institute)
ACF	-- actual cubic feet
ACFM	-- actual cubic feet per minute
ACRS	-- accelerated cost recovery system
ADA	-- anthraquinone disulfonic acid
ADR	-- asset depreciation range
AMB	-- ambient
BHP	-- brake horsepower
BOD	-- biochemical oxygen demand
BP	-- annual by-product credit
BPSD	-- barrels per stream day
CA	-- carbon adsorption
CC	-- total annual capital charge
COD	-- chemical oxygen demand
CPB	-- per-barrel control cost
CS/SS	-- carbon steel/stainless steel
DCF	-- discounted cash flow
DCF ROR	-- discounted cash flow rate of return
DDP	-- Detailed Development Plan
DEA	-- diethanolamine
DGA	-- diglycolamine
DIPA	-- diisopropanolamine
DOC	-- direct annual operating cost
DRI	-- Denver Research Institute
ED	-- electrodialysis
ESC	-- annual extra start-up costs
ESP	-- electrostatic precipitator
FCC	-- fixed capital cost

ABBREVIATIONS (cont.)

FGD -- flue gas desulfurization
FGR -- flue gas recirculation
fpm -- feet per minute
gpd/ft^2 -- gallons per day per square foot
gpm -- gallons per minute
gpt -- gallons per ton
HP -- horsepower
IBP -- initial boiling point
IOC -- indirect annual operating cost
ITC -- investment tax credit
LHV -- low heating value
LTPSD -- long tons per stream day
MDEA -- methyldiethanolamine
MEA -- monoethanolamine
MEB -- multiple effect boiling
meq -- milliequivalent
MIS -- Modified In Situ
MMBtu -- million British thermal units
MSF -- multistage flash
MTPSD -- metric tons per stream day
MW -- megawatts
MWt -- molecular weight
pcf -- pounds per cubic foot
PCTM -- Pollution Control Technical Manual
POM -- polynuclear organic matter
ppmv -- parts per million, by volume
ppmw -- parts per million, by weight
PSD -- Prevention of Significant Deterioration
psia -- pounds per square inch, absolute
psig -- pounds per square inch, gauge
RBOSC -- Rio Blanco Oil Shale Company
RF -- fixed capital charge rate
RHC -- reactive hydrocarbons

ABBREVIATIONS (cont.)

RO	-- reverse osmosis
RSH	-- alkyl thiols, mercaptans
RW	-- working capital charge rate
SCF	-- standard cubic foot
SCFM	-- standard cubic feet per minute
SCOT	-- Shell Claus Off-gas Treating
SCR	-- selective catalytic reduction
SEA	-- standard economic assumptions
SNPA/DEA	-- Societe Nationale des Petroles d'Aquitaine/ diethanolamine
SS	-- stainless steel
STC	-- annual severance tax credit
SWEC	-- Stone and Webster Engineering Corporation
TC	-- total annual control cost
TDS	-- total dissolved solids
TEG	-- triethylene glycol
TIA	-- annual property tax and insurance allowance
TOC	-- total annual operating cost
TPM	-- total particulate matter
TPSD	-- tons per stream day
TSS	-- total suspended solids
UF	-- ultrafiltration
USBM	-- U.S. Bureau of Mines
VCE	-- vapor compression evaporation
VOC	-- volatile organic compounds
WAO	-- wet air oxidation
WC	-- working capital
WPA	-- Water Purification Associates

CONVERSION FACTORS

1 pound, lb = 453.5924 grams, g
0.4536 kilograms, kg

1 ton = 0.9072 metic tons, tonnes

1 inch, in = 2.5400 centimeters, cm

1 foot, ft = 30.4800 centimeters, cm
0.3048 meters, m

1 mile, mi = 1,609.3440 meters, m
1.6093 kilometers, km

1 square inch, in^2 = 6.4516 square centimeters, cm^2

1 square foot, ft^2 = 0.0929 square meters, m^2

1 square mile, mi^2 = 2.5900 square kilometers, km^2
258.9988 hectares, ha

1 acre = 4,046.8564 square meters, m^2
0.4047 hectares, ha

1 cubic inch, in^3 = 16.3871 cubic centimeters, cm^3

1 cubic foot, ft^3 = 28.3161 liters, l

1 gallon, gal = 3.7853 liters, l
0.0038 cubic meters, m^3

1 barrel, bbl = 158.9828 liters, l
0.1590 cubic meters, m^3

1 acre-foot = 1,233.4818 cubic meters, m^3

1 pound/square inch, psi = 70.3070 grams/square centimeter, g/cm^2
0.0680 atmospheres, atm
0.3591 millimeters of mercury, mm of Hg

1 pound/cubic inch, lb/in^3 = 27.6799 grams/cubic centimeter, g/cm^3
27.6807 grams/milliliter, g/ml

CONVERSION FACTORS (cont.)

1 pound/cubic foot, pcf, lb/ft^3 = 0.0160 grams/cubic centimeter, g/cm^3
16.0185 kilograms/cubic meter, kg/m^3

1 gallon per ton, gpt = 4.1726 liter/tonne, l/tonne

1 barrel per day, BPD = 0.1590 cubic meters/day, m^3/d

1 gallon per minute, gpm = 0.0631 liters/second, l/s

1 British thermal unit, Btu = 251.9958 gram-calories, g-cal
1,054.1800 joules, J

1 million British thermal units, MMBtu = 292.8750 kilowatt-hours, kW-hr

1 British thermal unit/pound, Btu/lb = 0.5556 gram-calories/gram, g-cal/g

1 British thermal unit/cubic foot, Btu/ft^3 = 8.8994 gram-calories/liter, g-cal/l

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SECTION 1

INTRODUCTION

1.1 PURPOSE

Future U.S. energy production envisions the development of an environmentally acceptable, commercial synthetic fuels industry. As part of this overall effort, the Environmental Protection Agency (EPA), Office of Research and Development, has for the past several years undertaken extensive studies to determine synthetic fuel plant waste stream characteristics and potentially applicable pollution control systems.

The purpose of the Pollution Control Technical Manuals (PCTMs) is to convey, in a summarized and readily useful manner, information on synfuel waste stream characteristics and pollution control technology as obtained from studies by the EPA and others. The documents provide waste stream characterization data and describe a wide variety of pollution controls in terms of estimated performance, cost and reliability. The PCTMs contain no legally binding requirements, no regulatory guidance, and include no preference for process technologies or controls. Nothing within these documents relieves a facility from compliance with existing or future environmental regulations or permits.

The Pollution Control Technical Manuals consist of a set of seven discrete documents. There are six process specific PCTMs and a more general appendix volume which describes over fifty pollution control technologies. Application of pollution controls to a particular synfuel process is described in each process specific manual. The seven manuals are:

- *Pollution Control Technical Manual for Lurgi Based Indirect Coal Liquefaction and SNG*
- *Pollution Control Technical Manual for Koppers-Totzek Based Indirect Coal Liquefaction*
- *Pollution Control Technical Manual for Exxon Donor-Solvent Direct Coal Liquefaction*
- *Pollution Control Technical Manual for Lurgi Oil Shale Retorting with Open Pit Mining*
- *Pollution Control Technical Manual for Modified In Situ Oil Shale Retorting Combined with Lurgi Surface Retorting*

- *Pollution Control Technical Manual for TOSCO II Oil Shale Retorting with Underground Mining*
- *Control Technology Appendices for Pollution Control Technical Manuals*

By focusing on specific process technologies, the PCTMs attempt to be as definitive as possible on waste stream characteristics and control technology applications. This focus does not imply any EPA recommendations for particular process or control designs. Those described in the manuals are intended as representative examples of processes and control technologies that might be used. The design of the PCTMs, from process description through waste stream characterization and control technology evaluation, is intended to provide the user with a comprehensive understanding of the environmental factors inherent in operating synthetic fuel plants.

Control technology discussions presented in the PCTMs reflect pollutant removal levels which are believed to be achievable with currently available control technologies based upon existing data. Since there are no domestic commercial-scale synfuels facilities, the data base supporting this document was derived from bench- and pilot-scale synfuel facilities, developer's estimates, engineering analyses and analogue industries. As commercial synthetic fuel plants are built, the EPA will continue conducting research in order to develop a more comprehensive data base. Based on findings from these future studies, the Agency may update these documents or promulgate industry specific standards. In the interim, the Agency encourages facility planners, permit officials, and other interested parties to take advantage of the information contained in these documents.

1.2 APPROACH

The approach taken in developing this manual is to describe, in detail, an oil shale facility which has been proposed for development and to emphasize its pollution control aspects. This facility is the basis for the case study described in Section 2 "Summary of Study Features," Section 3 "Process Flow Diagrams and Flow Rates," and Section 4 "Inventory and Composition of Plant Process and Waste Streams." The process descriptions and control technologies presented in this case study are based on documents (identified in Section 1.3) published by the proposed facility developers and parallel, as closely as possible, the current thinking of the developers.

This manual examines Lurgi-Ruhrgas aboveground retorting with open pit mining and pit backfilling as proposed by Rio Blanco Oil Shale Company for development of Federal Lease Tract C-a in the Piceance Basin of Colorado. It should be noted, however, that effective August 1, 1982, Rio Blanco Oil Shale Company and the U.S. Department of the Interior (DOI) agreed to a suspension of operations and production for a period of 5 years or until DOI issues to Rio Blanco Oil Shale Company a lease for land other than Tract C-a (U.S. DOI, July 29, 1982). The company is seeking this additional land for purposes connected with operations on Tract C-a, including the disposal of processing wastes.

In order to enhance the flexibility of this manual, and since oil shale development plans are continually changing, Section 5 "Pollution Control Technology" expands beyond the case study description to examine other pollution control technologies and approaches that may be applicable to the waste sources identified in the case study. While controls applied to major gaseous, liquid, and solid streams described in the case study are those which have been proposed by the developer, Section 5 also examines alternative pollution control technologies on a stream-by-stream basis. For each stream receiving control, all major control technologies are discussed, while some example technologies are analyzed in considerable depth. Stream flow rates and pollutant characteristics are used in estimating the size, performance, and cost of the controls, and secondary streams resulting from the pollution control activities are identified.

It should be noted that the case study, as described in Sections 2, 3 and 4 of this manual, would begin approximately 30 years after the initial start of development operations on Tract C-a. Due to the space requirements for simultaneous production and backfilling operations, the first 30 years are spent developing the mine to a point when waste backfilling can commence without interfering with production. Also, water management and treatment activities would have reached a steady-state condition by this time. By examining the mining, backfilling, and water management/treatment operations under steady-state conditions, a more useful analysis can be made for these operations. This does not impair the usefulness of examining control technologies for the Lurgi retorting process since it would be operating under stable conditions from the outset.

1.3 DATA SOURCES

This manual focuses on the plans that have emerged over the past few years for the development of Federal Lease Tract C-a. The operation of the tract is monitored under the Federal Prototype Oil Shale Leasing Program through the U.S. Department of Interior's Minerals Management Service. Under this program, a Detailed Development Plan (DDP), modifications to the DDP, and extensive environmental information must be submitted on a regular basis to the Minerals Management Service by the lease operators. The DDP and subsequent modifications to the DDP submitted by the developers of Tract C-a were the principal data sources used to prepare the case study described in Sections 2, 3 and 4 of this manual.

The first commercial development plan, or DDP, for Tract C-a was published in 1976 (Gulf Oil Corp. and Standard Oil Co. [Indiana], March 1976). This plan called for open pit mining of 119,000 tons per stream day (TPSD) of raw shale to produce approximately 56,000 barrels per stream day (BPSD) of hydrotreated shale oil. According to the DDP, the processing facilities and waste disposal site were to be located off tract; however, the Department of Interior, based on acreage restrictions in the Minerals Leasing Act, denied the request for additional federal land. As a result, the developers submitted a revised DDP to the Area Oil Shale Office (now part of the Minerals Management Service) in 1977 (Gulf Oil Corp. and Standard Oil Co. [Indiana], May 1977). This modified plan was based on producing 76,000 BPSD of shale oil by using a combination of the Modified In Situ (MIS) and TOSCO II retorting technologies. The MIS development was planned to take place in a modular fashion, consisting of burning several small- and large-sized retorts over a period of ten years. In 1981, another modular program was incorporated into the DDP to demonstrate the feasibility of open pit mining and Lurgi-Ruhrgas aboveground retorting (Rio Blanco Oil Shale Co., February 1981).

The case study described in this manual is based on a combination of open pit mining from the original 1976 DDP and Lurgi surface retorting as described in the 1981 modification to the DDP. In addition, the 1977 revised DDP provided the basic site description and hydrologic data. Although these three documents were the major sources used in deriving the process, pollution control, and other information presented in Sections 2, 3 and 4 of this manual, several supplemental sources were also used and they are cited throughout the document.

Where available, actual data from the various scale operations in oil shale processing were used. It is believed that these data accurately reflect the major technical features which will be encountered in a commercial oil shale industry. In addition, technologies from analogue industries are transferred, when appropriate. When necessary, engineering analysis and judgment provided by the authors of this manual (Denver Research Institute, Stone and Webster Engineering Corporation and Water Purification Associates) and vendor information were used. In each case, all assumptions required to carry out the analyses are listed, and areas lacking hard data are identified (see Sections 1.5 and 7 for more detailed discussions).

1.4 STATE OF TECHNOLOGY DEVELOPMENT

As stated above, the processing plant considered in this manual is based on a combination of information from the three different DDPs submitted for Tract C-a. Approximately 119,000 TPSD of raw shale will be mined using open pit mining. In addition, 62,100 TPSD of overburden and 11,900 TPSD of subgrade shale (subore) will be removed. The raw shale on Tract C-a has an average oil yield of 23 gallons per ton (gpt) based on the modified Fischer assay. This shale will be processed in 13 Lurgi-Ruhrgas aboveground retorts, at an efficiency equal to 100% of Fischer assay, to eventually produce 63,140 BPSD of crude shale oil (The stream-day rates are the maximum, 24-hr/day rates that can be achieved; however, occasional equipment failure and required maintenance result in a reduced production efficiency. Normally, the plant can be expected to operate at 90% of its capacity on a long-term basis, or for 328.5 calendar days per year. Thus, the calendar-day production rates would be 90% of the stream-day rates.) The current status of the mining and retorting technologies is reviewed below.

1.4.1 Open Pit Mining

The open pit mine considered in this manual would be the largest of its type in the world. This mining method is used in other industries, such as copper ore mining, but the scale used is considerably smaller than that required in this study. Some preliminary design work for a commercial-scale open pit on Tract C-a was published in the original DDP, while the second, or revised, DDP did not consider the method at all. An experimental (small-size) open pit was subsequently described in the modification to the DDP, but this plan, which also included a full-scale Lurgi retort demonstration program, was suspended in late 1981 due to rising cost estimates for the demonstration program. Thus, only a limited amount of design and/or engineering data for an open pit mine on Tract C-a is available.

Although open pit mining is commonly used in other industries, the design for any pit should be evaluated on an individual basis, since its design and potential environmental impacts would be site dependent. For example, on Tract C-a the pit would intercept the upper and lower aquifers of the Piceance Creek Basin and, consequently, alter the hydrology of the area. Dewatering the strata in the vicinity of the pit would reduce the problems associated with its development, but several legal, engineering, and environmental issues may be raised and would need to be resolved. The excess groundwater would need to be reinjected, discharged, or disposed of in some manner. The processing and disposal facilities, if located on the tract, would take up a significant portion of the available land and this would severely hamper the development of the pit. Backfilling of the pit with the wastes could ease the space problem to some extent, but the logistics of having simultaneous mining and backfilling operations require an extensive effort. It was estimated in the original DDP that approximately 30 years of commercial-scale pit development would be necessary before the backfilling could be initiated.

1.4.2 Lurgi-Ruhrgas Technology

Small Lurgi-Ruhrgas pilot plants have been operated by Lurgi Kohle und Mineralötechnik GmbH in West Germany. The necessity to ship ore to West Germany has limited the amount of available test data. To date, the experience relevant to this manual is limited to three tests: one in 1976 on shale from the Colony mine in Colorado, and two in 1980 on shales from Tracts C-a and C-b. The earlier (1976) test was run in a 5-ton/day pilot plant, while the 1980 tests were performed in a 25-ton/day plant. Data from the 1980 test on Tract C-a shale, published in the modified DDP for the Lurgi demonstration module, were used in this manual (Rio Blanco Oil Shale Co., February 1981). Tests have been run on other shales and reported in the literature (Marnell, September 1976; Schmalfeld, July 1975), but substantially different results were obtained.

The Lurgi-Ruhrgas demonstration plant processing 4,400 TPSD of shale on Tract C-a (as proposed in the modified DDP) was to be operational by early 1983, but these plans were suspended indefinitely during the summer of 1981 in favor of building and operating a 5-ton/day pilot plant at Gulf Oil's research facility in Pennsylvania.

1.5 ASSUMPTIONS

In performing a detailed analysis of the Lurgi-Ruhrgas retorting process used with open pit mining, a number of specific assumptions which influence the results of the analysis and their interpretation were made. The underlying, major assumptions relating to pollution control performance and cost, as well as the bases behind the assumptions, are summarized in this section.

1.5.1 Pollution Control and Performance Estimates

In the process of preparing this manual, applicable pollution control technologies for different waste streams were reviewed, and controls proposed by industry were evaluated to the point that performance and cost could be estimated. Equipment vendors' estimates and guarantees were used whenever available. Other performance levels were estimated using laboratory testing data. These performance estimates should be viewed tentatively because very little data based on actual source testing exist. The major pollution controls evaluated in this manual are presented in Table 1.5-1, along with the performance levels estimated as a result of the analysis.

The major air pollution control technologies evaluated (electrostatic precipitator, Stretford) are commercially available and are used in other industries at a scale similar to that involved in this manual; therefore, operational difficulties in adapting these technologies to oil shale processing are not expected to be great and may primarily involve adapting these technologies to oil shale off-gas characteristics.

In the area of water pollution control, it has been proposed by the developer that the plant will achieve zero-discharge of the process generated waters, but that excess mine water will need to be discharged. The process waters are treated to the degree necessary for reuse. The technologies considered (ammonia recovery, aeration pond) have been used in analogue industries and can be expected to be employed successfully in the oil shale industry. Waters used in auxiliary plant operations are also treated since the wastes produced from these operations may be used in processed shale moistening and thus may become a source of pollution. Reuse of some waters may negate a need for pollution control; in such cases, no pollution controls in a conventional sense are applied.

Solid wastes are managed by backfilling the open pit. This approach was mentioned in the original DDP for Tract C-a and was to commence after 30 years of pit development and off-site disposal of the wastes, but detailed plans were not presented. The pollution control technologies that are judged appropriate for open pit backfilling include surface hydrology technologies, such as a runoff collection system and pumps during the project life, subsurface hydrology technologies primarily involving the monitoring of the groundwater, and surface stabilization technologies for dust suppression, revegetation, etc. These technologies are traditional practices associated with solid waste management in other industries.

1.5.2 Components of Pollution Control Cost Estimates

Fixed capital and direct annual operating costs were estimated for each piece of pollution control equipment and each control activity. These figures were then used, along with economic assumptions, to calculate total annual control costs which include an annual charge for capital. The total annual capital charge provides for a required after-tax return on investment of 12 percent. The approach used to estimate the capital and operating costs and the economic analysis techniques applied to these data are summarized in Tables 1.5-2 through 1.5-4.

TABLE 1.5-1. PERFORMANCE LEVELS ESTIMATED FOR MAJOR POLLUTION CONTROLS

Control Description	Pollutant Controlled	Control Level Estimated
AIR POLLUTION CONTROL		
Baghouses	Raw and Processed Shale Dust	99.7%[a]
Water Sprays	Raw and Processed Shale Dust	50%[b]
Foam Sprays	Raw and Processed Shale Dust	85%[b]
Electrostatic Precipitator	Processed Shale Dust	99.9%[c]
Holmes-Stretford Gas Treating Process	H_2S	30 ppmv[d]
WATER POLLUTION CONTROL		
Ammonia Recovery Unit	NH_3	99%[e]
	Organic Matter	25% reduction[e]
Aeration Pond	COD	25%[f]
SOLID WASTE MANAGEMENT		
Mine Backfill	Processed Shale, Sludges, Blowdowns, Concentrates, etc.	N/A

Sources:

[a] Vendor estimates assuming 10 grains/ACF inlet loading.

[b] SWEC estimates. Efficiency is based on the quantity of airborne material. The efficiency may be higher in terms of the material contacted by the sprays.

[c] Rio Blanco Oil Shale Co., February 1981.

[d] Based on Peabody Process Systems, Inc., February 1981.

[e] Estimates from treatability studies on similar waters conducted by Water Purification Associates, unpublished.

[f] WPA estimates.

Fixed capital and direct annual operating cost estimates were developed on a component basis, using current cost data from the actual installation and operation of similar facilities and using vendor quotes for major equipment items. Capital cost estimates are expected to have an average accuracy of ±30 percent. This level of accuracy can only be verified by actual equipment installation. Experience in using the cost estimating procedures for units which actually have been constructed and operated indicate that this level of accuracy should be achievable if the unit installed is exactly as described in determining the cost estimate. Any design changes could cause the actual installed capital cost to fall outside the range.

Table 1.5-2 lists the components estimated in determining the installed fixed capital cost of pollution control equipment. For simple equipment, all components may not be present. For large and complex equipment, estimating the cost of each component may be a major effort. A description of the major equipment included in each capital cost estimate is provided in Section 5 of this document.

TABLE 1.5-2. COMPONENTS OF FIXED CAPITAL COST ESTIMATES

Components
Major Equipment (vendor quotes)
Site Preparation, Excavation and Foundations
Concrete and Rebar
Support Structures
Piping, Ductwork, Joints, Valves, Dampers, etc.
Duct and Pipe Insulation
Pumps and Blowers
Electrical
Instrumentation and Controls
Monitoring Equipment
Erection and Commissioning
Painting
Buildings
Engineering and Other Indirect Costs
Contractor's Fee
Contingency Allowances

Source: DRI.

Table 1.5-3 shows the components comprising direct annual operating costs. Operating supplies include such items as baghouse bags. Maintenance includes the cost of parts used, but the needed inventory of replacement parts is included in fixed capital cost. The cost of water consumed is not included due to uncertainties in estimating the value of water. Direct annual operating costs do not include by-product credits; however, by-product credits are included in total annual operating costs. The operating costs (direct, indirect and total) for each pollution control, along with a detailed discussion of how the costs were determined, are presented on a component basis in Section 6.

TABLE 1.5-3. COMPONENTS OF DIRECT ANNUAL OPERATING COST ESTIMATES

Components
Maintenance and Maintenance Supplies
Operating Supplies
Operating Labor
Cooling Water
Steam
Electricity
Fuel
Indirect Costs (e.g., supervision, laboratory, etc.)*

* Indirect costs are included in the labor rate.

Source: DRI.

Table 1.5-4 presents the major economic assumptions used in the cost evaluations. Most economic assumptions have been standardized so that the results found in all of the oil shale PCTMs may be compared. A sensitivity analysis was performed (see Section 6 "Pollution Control Costs") to determine the effects of changes in some of the standard economic assumptions. These changes include delayed start-up, changing capital and operating costs, financing considerations and others. All of the oil shale PCTMs use a discounted cash flow approach (DCF) and constant dollars (mid-1980).

TABLE 1.5-4. SUMMARY OF MAJOR STANDARD ECONOMIC ASSUMPTIONS USED IN CONTROL COST EVALUATIONS

Assumptions[a]
• Approach: Discounted Cash Flow Evaluation (DCF)
• Method: Revenue Requirement determined from capital charge plus operating cost[b]
• Required DCF ROR: 12% (100% equity basis)
• Cost Base: Mid-1980 constant dollars
• Income Tax: In accordance with current regulations (48% combined tax rate, 20% investment tax credit); tax credits and allowances can be passed through to a parent company that can benefit from them immediately, without waiting for the project to become profitable
• Project Timing: 4 years construction, 20 years life
• Normal Plant Output: 63,140 barrels per stream day (net, after in-plant use)
• Operating Factors: Year 1 - 50% Year 2 - 75% Years 3-20 - 90%

[a] A more detailed list of assumptions is presented in Section 6, Table 6.2-1.

[b] This method permits accurate costs to be determined separately for each control using the DCF approach, without the need for an estimate of total plant cost.

Source: DRI.

1.6 UNIQUE FEATURES

Three oil shale retorting processes were selected for the oil shale PCTMs to allow consideration of different types of retorting processes, mining and disposal techniques, and pollution control technologies. Some features are found in more than one manual, but each process examined has important unique features which are listed in Table 1.6-1.

Table 1.6-2 lists the pollution control technologies examined in the three PCTMs. The table is designed to assist the reader in locating detailed information on any specific control technology.

TABLE 1.6-1. MAJOR FEATURES OF THE OIL SHALE PCTMs

Feature	PCTMs		
	TOSCO II	MIS-Lurgi	Lurgi-Open Pit
MINING			
Underground Room-and-Pillar	X		
Underground MIS		X	
Open Pit			X
RETORTING			
Aboveground	X	X	X
Underground		X	
Direct-heated		X	
Indirect-heated	X	X	X
Solid-to-Solid Heat Transfer	X	X	X
Gas-to-Solid Heat Transfer		X	
Resource Recovery from Processed Shale		X	X
High Carbon Processed Shale	X		
Low Carbon Processed Shale		X	X
Raw Shale Preheating	X		
PROCESSING			
High Btu Off-gas	X	X	X
Low Btu Off-gas		X	
Oil Fractionation	X	X	X

(Continued)

TABLE 1.6-1 (cont.)

Feature	TOSCO II	MIS-Lurgi	Lurgi-Open Pit
PROCESSING (cont.)			
Oil Upgrading	X		
Gas Upgrading (for sale)	X		X
In-Plant Fuel Use	X	X	
Excess Electricity		X	
POLLUTION CONTROL			
Retort Gas Cleanup	X	X	X
Process Water Cleanup	X	X	X
Excess Water Discharge			X
By-product Recovery	X	X	X
WASTE DISPOSAL			
Surface Landfill	X	X	
Permitted Design	X		
Open Pit Backfill			X
Groundwater Contamination Potential (subsurface disposal or retort abandonment)		X	X
Surface Water Contamination Potential (valley fill)	X	X	

Note: the TOSCO II, MIS-Lurgi and Lurgi-Open Pit columns are printed under a spanning header "PCTMs".

Source: DRI.

TABLE 1.6-2. POLLUTION CONTROL TECHNOLOGIES EXAMINED IN THE OIL SHALE PCTMs

	PCTMs		
Control Technology	TOSCO II	MIS-Lurgi	Lurgi-Open Pit
AIR POLLUTION			
Diethanolamine (DEA)	X		
Methyldiethanolamine (MDEA)	X		
Claus	X		
Wellman-Lord	X		
Stretford	X	X	X
Shell Claus Off-gas Treating (SCOT)	X		
Limestone Scrubber (FGD)		X	
Absorber/Cooler		X	
Low Flare	X		
High Energy Venturi Wet Scrubber	X		
Venturi Wet Scrubber	X	X	
Electrostatic Precipitator		X	X
Thermal Oxidizer	X		
Fabric Filter (baghouse)	X	X	X
Foam Sprays	X	X	X
Water Sprays	X	X	X
Double Seal Oil Storage	X	X	X
Refrigerated Ammonia Storage	X	X	X
Catalytic Converter	X	X	X
Maintenance	X	X	X

(Continued)

TABLE 1.6-2 (cont.)

	PCTMs		
Control Technology	TOSCO II	MIS-Lurgi	Lurgi-Open Pit
WATER MANAGEMENT			
Ammonia Recovery	X	X	X
Biological Oxidation	X		
Steam Stripper	X	X	
Kettle Evaporator		X	
Reverse Osmosis	X	X	X
Carbon Adsorption	X	X	X
Wet Air Oxidation	X	X	
Vapor Compression Evaporation	X		
Reinjection			X
Multimedia Gravity Filtration		X	
Clarifier	X	X	X
Process Oil/Water Separator	X	X	X
Runoff Oil/Water Separator	X	X	X
Boiler Feedwater Treatment	X	X	X
Cooling Tower Makeup Treatment	X	X	X
Equalization Pond	X	X	X
Aerated Pond			X
Solar Pond	X	X	X

(Continued)

TABLE 1.6-2 (cont.)

Control Technology	PCTMs		
	TOSCO II	MIS-Lurgi	Lurgi-Open Pit
SOLID WASTE MANAGEMENT			
Runoff Collection System	X	X	X
Upper Embankments	X		
Lower Embankments	X	X	
Runon Collection System	X	X	
Stilling Basin	X		
Water Impoundment	X		
Leachate Collection System	X	X	
Spring Collection/Underdrains	X	X	
Covers and Bottom Liners	X	X	X
MIS Spent Retort Treatment		X	
Dust Supression	X	X	X
Surface Reclamation	X	X	X
Piezometers			X

Source: DRI.

1.7 ORGANIZATION AND USE OF THE MANUAL

Following this "Introduction" to the PCTM are six major sections which present material ranging from basic background information to detailed pollution control data and costs. In addition, a complete listing of all information sources used to develop the manual is provided in Section 8 "References." A brief description of each of the major sections is presented below.

Section 2 provides an overview of the Lurgi retorting process and the case study examined in the manual. It gives background information on the proposed project development, including the site involved, retorting and mining processes, and the pollution controls proposed by the developers of Tract C-a.

Section 3 expands upon the case study outlined in Section 2. Detailed process flow diagrams and descriptions are given for each unit process. Individual streams, their mass flow rates, and other characteristics are generated during the unit process analyses, and this information is the basis for detailed stream discussions presented in Section 4.

Section 4 provides the detailed compositions for the major process streams identified in Section 3. These parameters are then used in designing and costing the pollution control technologies discussed in Section 5. All streams identified in Section 3 are inventoried by media (gas, liquid, solid) and important features of each stream are noted (Tables 4.1-1, 4.1-3 and 4.1-5, respectively). Also, the detailed stream compositions are summarized by media (Table 4.1-2, 4.1-4 and 4.1-6).

Section 5 presents concise inventories of the available control technologies and approaches for air, water and solid wastes. Key features of each technology are briefly described and many leading technologies are analyzed in greater depth. The fixed capital and direct annual operating costs and design details for the leading technologies are also presented.

Section 6 presents the total annual and per-barrel cost of pollution control based upon the cost data developed for the control technologies in Section 5 and the standard economic assumptions used in all oil shale PCTMs. This section also analyzes the sensitivity of the control costs to variations in the standard economic assumptions and capital and operating cost parameters.

Section 7 discusses the limitations of the data base used in the preparation of the manual. It also identifies important areas that may require more research.

Table 1.7-1 provides a composite list of the major process and waste streams generated by the facility described in Section 3. All streams are identified with a unique name and number. An asterisk (*) is placed next to the stream number if the stream comes into contact with the environment at any point in the process, and a descriptive letter is given to identify the state of the stream, i.e., gaseous (G), liquid (L) or solid (S). Also, cross-references are included for the flow diagrams in which the stream is produced and/or processed (Section 3), detailed composition tables (Section 4), and applicable control technologies (Section 5) to allow tracking of the stream from its origin to its final disposition.

For example, stream 34 in Table 1.7-1 is the retort gas--a gaseous stream that does not contact the environment. It is produced by processing of the retort vapors (stream 26) from the Lurgi retorts, as illustrated in Figure 3.3-4, Section 3. Table 4.2-12 (Section 4) provides the detailed composition of the gas, and Section 5.1.3 ("Nitrogen Oxides Control") briefly discusses approaches to lower the ammonia content of the gas in order to

TABLE 1.7-1. COMPOSITE LIST OF STREAMS

Stream Number	Description of Stream	Type of Stream	Cross-References		
			Flow Diagram Numbers	Composition Table Numbers	Control Technology Sections
1*	Raw Shale Feed	S	3.3-2, 3.3-3	4.2-2	--
2*	Subore	S	3.3-2, 3.3-10	--	5.3.1, 5.3.2, 5.3.3, 5.3.4
3*	Overburden	S	3.3-2, 3.3-10	--	5.3.1, 5.3.2, 5.3.3, 5.3.4
4*	Mine Water	L	3.3-2, 3.3-11	4.2-22	5.2.1, 5.2.3
5*	Primary Crusher (ore), Baghouse Emission	G	3.3-2	--	5.1.1
6*	Primary Crusher (subore), Baghouse Emission	G	3.3-2	--	5.1.1
7*	Primary Crusher (overburden), Baghouse Emission	G	3.3-2	--	5.1.1
8*	Raw Shale Conveyor Transfer Point, Baghouse Emission	G	3.3-2	--	5.1.1
9*	Swinging Boom Stacker, Baghouse Emission	G	3.3-2	--	5.1.1
10*	Coarse Ore Conveyor Transfer Point, Baghouse Emission	G	3.3-2	--	5.1.1
11*	Secondary Crusher, Baghouse Emission	G	3.3-2	--	5.1.1
12*	Secondary Crushing to Screening Conveyor Transfer Point, Baghouse Emission	G	3.3-2	--	5.1.1
13*	Secondary Screening, Baghouse Emission	G	3.3-2	--	5.1.1
14*	Secondary Screening Conveyor Transfer Point, Baghouse Emission	G	3.3-2	--	5.1.1
15*	Tertiary Crusher, Baghouse Emission	G	3.3-2	--	5.1.1
16*	Tertiary Crushing to Tertiary Screening Conveyor Transfer Point, Baghouse Emission	G	3.3-2	--	5.1.1
17*	Tertiary Screening, Baghouse Emission	G	3.3-2	--	5.1.1
18*	Tertiary Screening to Fine Ore Storage Conveyor Transfer Point, Baghouse Emission	G	3.3-2	--	5.1.1
19*	Fine Ore Storage, Baghouse Emission	G	3.3-2	--	5.1.1

(Continued)

TABLE 1.7-1 (cont.)

Stream Number	Description of Stream	Type of Stream	Cross-References Flow Diagram Numbers	Composition Table Numbers	Control Technology Sections
20*	Retort Feed Hopper Conveyor Transfer Point, Baghouse Emission	G	3.3-2	--	5.1.1
21*	Retort Feed Hopper, Baghouse Emission	G	3.3-2	--	5.1.1
22*	Baghouse Dusts	S	3.3-2, 3.3-3	4.2-2	--
23*	Processed Shale Load-out Hopper, Baghouse Emission	G	3.3-2, 3.3-10	--	5.1.1
24*	Diesel Emissions	G	3.3-2, 3.3-10	--	5.1.4, 5.1.5
25	Combustion Air to Lift Pipes	G	3.3-3	--	--
26	Retort Vapors	G	3.3-3, 3.3-4	4.2-9	--
27	High Pressure Steam	G	3.3-3, 3.3-6	--	--
28*	Blowdown from Waste Heat Boiler	L	3.3-3, 3.3-11	--	5.2.1
29*	Processed Shale	S	3.3-2, 3.3-3, 3.3-10	4.2-5, 4.2-6, 4.2-7	5.3.1, 5.3.2, 5.3.3, 5.3.4
30	Steam to Humidifier	G	3.3-3	--	--
31*	Lurgi Flue Gas	G	3.3-3	4.2-19	5.1.1
32*	Raw Shale Retort Feed Conveyor, Baghouse Emission	G	3.3-3	--	5.1.1
33	Raw Retort Gas	G	3.3-4	4.2-11	--
34	Retort Gas	G	3.3-4, 3.3-5	4.2-12	5.1.3
35	Retort Gas to Lift Pipes	G	3.3-3, 3.3-4	--	--
36	Light Oils to Storage	L	3.3-4	4.2-10	--
37	Light Oil Makeup to Naphtha Recovery	L	3.3-4, 3.3-5	--	--
38	Middle Oils to Storage	L	3.3-4	4.2-10	--
39	Diesel Fuel - Mining Equipment	L	3.3-2, 3.3-4, 3.3-10	--	--
40	Diesel Fuel - Disposal Equipment	L	3.3-4, 3.3-10	--	--
41	Gas Liquor	L	3.3-4, 3.3-9	4.2-20	5.2.1, 5.2.2, 5.2.3
42	Heavy Oils to Storage	L	3.3-4	4.2-10	--
43	Oily Dust	S	3.3-3, 3.3-4	--	--
44*	Fugitive Hydrocarbon Emissions from Storage Tanks	G	3.3-4	--	5.1.4
45	Naphtha-free Retort Gas	G	3.3-5, 3.3-6	4.2-13	5.1.2, 5.1.3

(Continued)

TABLE 1.7-1 (cont.)

Stream Number	Description of Stream	Type of Stream	Cross-References Flow Diagram Numbers	Composition Table Numbers	Control Technology Sections
46	Naphtha Product to Storage	L	3.3-5	4.2-14	--
47*	Hydrocarbon Emissions from Naphtha Storage	G	3.3-5	--	5.1.4
48	Compressed Naphtha-free Gas	G	3.3-6, 3.3-7	4.2-15	5.1.2, 5.1.3
49	Compressor Condensate	L	3.3-6, 3.3-9	4.2-20	5.2.1, 5.2.2, 5.2.3
50	Steam to Naphtha Recovery	G	3.3-5, 3.3-6	--	--
51	Steam to DEA Unit	G	3.3-6, 3.3-7	--	--
52	Steam to Stretford Unit	G	3.3-6, 3.3-8	--	--
53	Steam to Ammonia Recovery Unit	G	3.3-6, 3.3-9	--	--
54	Amine Makeup	L	3.3-7	--	--
55	TEG Makeup	L	3.3-7	--	--
56	Sweet Gas from DEA Unit	G	3.3-7	4.2-16	5.1.2
57	Dried Fuel Gas to Pipeline	G	3.3-7	4.2-17	--
58	Acid Gases from DEA Regeneration	G	3.3-7, 3.3-8	4.2-18	5.1.2
59*	Spent Amine	L	3.3-7, 3.3-10	--	5.3.1
60*	TEG Regeneration Vent Emission	G	3.3-7	--	--
61	Stripping Air to Stretford	G	3.3-8	4.2-18	--
62	Stretford Chemicals	L	3.3-8	--	--
63*	Stretford Treated Acid Gases	G	3.3-8	4.2-18	5.1.2
64	Stretford Oxidizer Vent Gas	G	3.3-3, 3.3-8	4.2-18	--
65*	Stretford Spent Liquor to Reclaim	L	3.3-8	--	--
66	Liquid Sulfur Product to Storage	L	3.3-8	4.2-18	--
67	Phosphoric Acid	L	3.3-9	--	--
68	Caustic (NaOH)	S	3.3-9	--	--
69	Steam Condensate from Ammonia Recovery	L	3.3-9, 3.3-11	--	--
70*	Stripped Gas Liquor	L	3.3-9, 3.3-11	4.2-21	5.2.3
71	Anhydrous Ammonia to Storage	L	3.3-9	4.2-21	5.1.3
72	Ammonia Overhead Vapors	G	3.3-3, 3.3-9	4.2-21	5.1.2, 5.2.2
73*	Processed Shale Conveyor Transfer Point, Baghouse Emission	G	3.3-10	--	5.1.1

(Continued)

TABLE 1.7-1 (cont.)

Stream Number	Description of Stream	Type of Stream	Cross-References: Flow Diagram Numbers	Cross-References: Composition Table Numbers	Cross-References: Control Technology Sections
74*	Fugitive Dusts	G	3.3-2, 3.3-10	--	5.1.1
75*	Excess Mine Water to Aeration Pond	L	3.3-11	4.2-23	5.2.3, 5.2.4
76*	Aerated Water to Discharge	L	3.3-11	4.2-23	--
77	Feedwater to Waste Heat Boiler	L	3.3-3, 3.3-11	--	--
78*	Total Processed Shale Moistening Water	L	3.3-3, 3.3-11	--	5.2.1
79	Cooling Water to Lurgi Oil Recovery	L	3.3-4, 3.3-11	--	5.2.3
80	Cooling Water to Naphtha Recovery	L	3.3-5, 3.3-11	--	5.2.3
81	Steam Condensate from Naphtha Stripper	L	3.3-5, 3.3-11	--	--
82	Cooling Water to Compression Cooling	L	3.3-6, 3.3-11	--	5.2.3
83	Cooling Water to DEA-TEG Treatment	L	3.3-7, 3.3-11	--	5.2.3
84	Steam Condensate from DEA-TEG Treatment	L	3.3-7, 3.3-11	--	--
85	Steam Condensate from Stretford	L	3.3-8, 3.3-11	--	--
86	Cooling Water Makeup to Stretford	L	3.3-8, 3.3-11	--	5.2.3
87	Process Water Makeup to Stretford	L	3.3-8, 3.3-11	--	5.2.1, 5.2.3
88*	Humidified Air Cooler Blowdown	L	3.3-4, 3.3-8, 3.3-11	--	5.2.1
89	Cooling Water to Ammonia Recovery	L	3.3-9, 3.3-11	--	5.2.3
90*	Water for Dust Palliatives	L	3.3-2, 3.3-10, 3.3-11	--	5.2.1, 5.3.4
91*	Processed Shale Revegetation Water	L	3.3-10, 3.3-11	--	5.2.1, 5.3.4
92*	Raw Shale Leachate	L	3.3-2, 3.3-11	4.2-3, 4.2-4	5.2.1, 5.3.3
93*	Storm Runoff	L	3.3-11	--	5.2.1, 5.3.2
94	Boiler Feedwater Makeup	L	3.3-11	--	5.2.3
95*	Service and Fire Water	L	3.3-11	--	5.2.1
96*	Mine Water Clarifier Sludge	L	3.3-11	--	5.2.1

(Continued)

TABLE 1.7-1 (cont.)

Stream Number	Description of Stream	Type of Stream	Cross-References		
			Flow Diagram Numbers	Composition Table Numbers	Control Technology Sections
97	Water to Cooling Tower Makeup Treatment	L	3.3-11	--	5.2.3
98	Treated Water to Cooling Tower	L	3.3-11	--	5.2.3
99*	Potable/Sanitary Water	L	3.3-11	--	5.2.1
100*	Water Evaporation from Mine Water Clarifier	G	3.3-11	--	5.2.3
101*	Used Sanitary Water to Municipal Treatment	L	3.3-11	--	5.2.1
102*	Treated Sanitary Water	L	3.3-11	--	5.2.1
103*	Sanitary Water Treatment Sludge	L	3.3-10, 3.3-11	--	5.3.4
104*	Boiler Feedwater Treatment Concentrate	L	3.3-11	--	5.2.1, 5.2.3
105*	Cooling Tower Blowdown	L	3.3-11	--	5.2.1, 5.2.3
106*	Cooling Tower Drift	L	3.3-11	--	5.2.3
107*	Cooling Tower Evaporation	G	3.3-11	--	5.2.3
108*	Equalization Pond Discharge to Processed Shale Moistening	L	3.3-11	--	5.3.3, 5.3.4
109*	Clarified Mine Water to Processed Shale Moistening	L	3.3-11	--	5.2.1, 5.3.3, 5.3.4
110*	Water Evaporation from Equalization Pond	G	3.3-11	--	5.2.1
111*	Aerated Pond Sludge	L	3.3-10, 3.3-11	--	5.2.3
112*	Miscellaneous HC Emission	G	--	--	5.1.4

* Indicates streams that come into contact with the environment.

Source: DRI.

control the nitrogen oxides emissions. Figure 3.3-4 indicates that Figure 3.3-5 is the destination for the gas.

Figure 3.3-5 exemplifies the processing of the retort gas to produce the naphtha-free retort gas (stream 45) for which the detailed composition is given in Table 4.2-13. The naphtha-free gas can be followed sequentially through Figures 3.3-6 (stream 48), 3.3-7 (stream 58) and 3.3-8 (stream 63) to illustrate the compression of the gas, removal of the acidic components from the gas, and release of the treated acid gases (after the removal and recovery of H_2S) into the atmosphere, respectively. Other process and waste streams can be followed in a similar manner.

SECTION 2

SUMMARY OF STUDY FEATURES

The Federal Prototype Oil Shale Leasing Program was initiated by the U.S. Department of the Interior (DOI) in 1974. The purpose of the program was to encourage commercial development of the energy resource in the Green River Oil Shale Formation. Six lease tracts, two each in the states of Colorado, Utah and Wyoming, were created and offered to the public on the basis of high bid. The first of these tracts, Federal Lease Tract C-a in Colorado, was subsequently awarded to Gulf Oil Corporation and Standard Oil Company (Indiana) after submission of a joint bonus bid of approximately $210 million. The two companies then formed a general partnership in 1978 and created the Rio Blanco Oil Shale Company (RBOSC) to operate the tract.

Under the requirements of the Leasing Program, Gulf and Standard submitted a Detailed Development Plan (DDP) to develop the tract using open pit mining and a combination of TOSCO II and gas combustion type aboveground retorting, with the understanding that off-tract disposal of the overburden and processed shale would be allowed so that the tract could be explored to its full potential (Gulf Oil Corp. and Standard Oil Co. [Indiana], March 1976). However, due to acreage restrictions in the Minerals Leasing Act, the DOI refused to grant additional Federal land for disposal purposes. Subsequently, a revised DDP was submitted emphasizing Modified In Situ (MIS) retorting with a supporting TOSCO II aboveground retorting facility (Gulf Oil Corp. and Standard Oil Co. [Indiana], May 1977). This plan did not require open pit mining or off-site disposal of the solid wastes.

In 1981, a modification was added to the DDP to demonstrate the feasibility of open pit mining with Lurgi-Ruhrgas aboveground retorting. Specifically, the plan called for an experimental open pit to support a Lurgi demonstration plant that would process approximately 4,400 TPSD of shale (Rio Blanco Oil Shale Co., February 1981). The retort was scheduled for operation in early 1983; however, due to rising cost estimates for the plant, the demonstration project was suspended in the summer of 1981. Recently, DOI approved a suspension of development operations on Tract C-a (U.S. DOI, July 29, 1982).

Currently, RBOSC is planning to build a Lurgi pilot plant (1 to 5 TPSD) at Gulf's research facility in Pennsylvania. The objective of the study is to obtain essential technical details on the Lurgi retorting technology.

This manual examines open pit mining as proposed in the original DDP of March 1976, combined with Lurgi retorting as proposed in the February 1981 modification to the DDP.

2.1 PROCESS OVERVIEW

The starting point for the analysis is some 30 years after the operation has reached full production capacity. This amount of time was estimated to be necessary to create enough working space in the pit so that mine backfilling could commence (Gulf Oil Corp. and Standard Oil Co. [Indiana], March 1976). The overburden and processed shale generated during these 30 years will be disposed of permanently on an off-tract location. In order to be consistent with other oil shale Pollution Control Technical Manuals, a project life of 20 years, following the initial 30 years, has been assumed for costing purposes. The wastes generated during this 20-year period are placed back in the pit. The dewatering of the aquifers is continued throughout the project, and the excess aquifer water is treated for surface discharge. The shale oil is not upgraded on site, but the retort gas is cleaned to pipeline quality so that it can be sold. Also, there is a potential for generating some electricity from the excess steam.

Open pit mining of 119,000 TPSD of raw oil shale with 62,100 TPSD of overburden and 11,900 TPSD of subgrade ore will be required for the commercial operation. The mining of approximately 193,000 TPSD of solids will be the largest open pit operation in the world; by comparison, the largest open pit at present is the Kennecott copper mine in Utah, which produces 110,000 TPSD of copper ore.

A full-sized Lurgi retort can process about 4,500 TPSD of oil shale. However, some of the noncritical units of the module, such as the feed hopper, collecting bin, and surge vessel, can be increased in size to accommodate two each of the critical units, such as the screw mixer, lift pipe, etc. As a result, approximately 9,150 TPSD of the shale can be processed in a single train on a 24-hr/day basis. Thirteen larger capacity trains would be required to process 119,000 TPSD of oil shale for the commercial operation.

A gross oil production rate of 65,167 BPSD (including naphtha) may be expected; this is based on a yield of 23 gallons of crude oil per ton of shale and a retorting efficiency of 100% of the modified Fischer assay. However, approximately 2,000 BPSD of naphtha are calculated to be consumed with the retort gas, which is used as supplemental fuel to balance the energy needs of the lift pipes; thus, the net oil production rate is 63,140 BPSD. The net retort gas production is at 149×10^3 lb/hr before naphtha removal and 122×10^3 lb/hr after recovering the naphtha. The gas rate to the pipeline is 62×10^3 lb/hr after cleanup. Approximately 600 gpm of process water, or gas liquor, are also produced, from which 22.6 TPSD of ammonia are recovered. The processed shale is produced and disposed of at a rate of 95,000 TPSD (dry basis).

The quantities defining the dimensions of the plant complex are listed in Table 2.1-1. Process related quantities have been estimated primarily from the data published by the developer (Gulf Oil Corp. and Standard Oil Co. [Indiana], March 1976; Rio Blanco Oil Shale Co., February 1981). These quantities form the basis for the technical analyses and discussions presented in this document.

TABLE 2.1-1. MAJOR PARAMETERS DEFINING THE SIZE OF THE COMMERCIAL PLANT COMPLEX

Parameter, Unit	Quantity
Net Oil Produced, BPSD	63,140*
Retorting Oil Yield, % Fischer Assay	100
Net Retort Gas Produced, 10^3 lb/hr	149
Treated Gas to Pipeline, 10^3 lb/hr (10^9 Btu/hr)	62 (1.2)
Total Solids Mined, TPSD	193,000
Raw Oil Shale, TPSD	119,000
Raw Shale Grade, gpt	23
Overburden, TPSD	62,100
Subore, TPSD	11,900
Processed Shale Disposed (dry basis), TPSD	94,956
Processed Shale Moistening Water, gpm	3,644
Flue Gas Produced, 10^3 lb/hr	7,195
Gas Liquor Produced, gpm	586
Sulfur Produced, MTPSD	7.6
Ammonia Produced, TPSD	22.6
Mine Water Produced, gpm	16,500
Mine Water Consumed, gpm (bbl/bbl of oil)	8,170 (4.4)
Mine Water Discharged, gpm	8,330
Number of Retort Trains	13
On-stream Factor, %	90
Project Duration, years	20
Total Land Area, acres	5,100
Open Pit Area, acres	1,150
Open Pit Surface Diameter, feet	7,900
Mine Depth, feet	1,350
Processed Shale Disposal Area, acres	1,150

* The gross oil production rate is 65,167 BPSD. Approximately 2,000 BPSD of the naphtha oil are used in the lift pipes.

Source: DRI estimates based on data from Gulf Oil Corp. and Standard Oil Co. (Indiana), March 1976, and Rio Blanco Oil Shale Co., February 1981.

2.1.1 Site Description

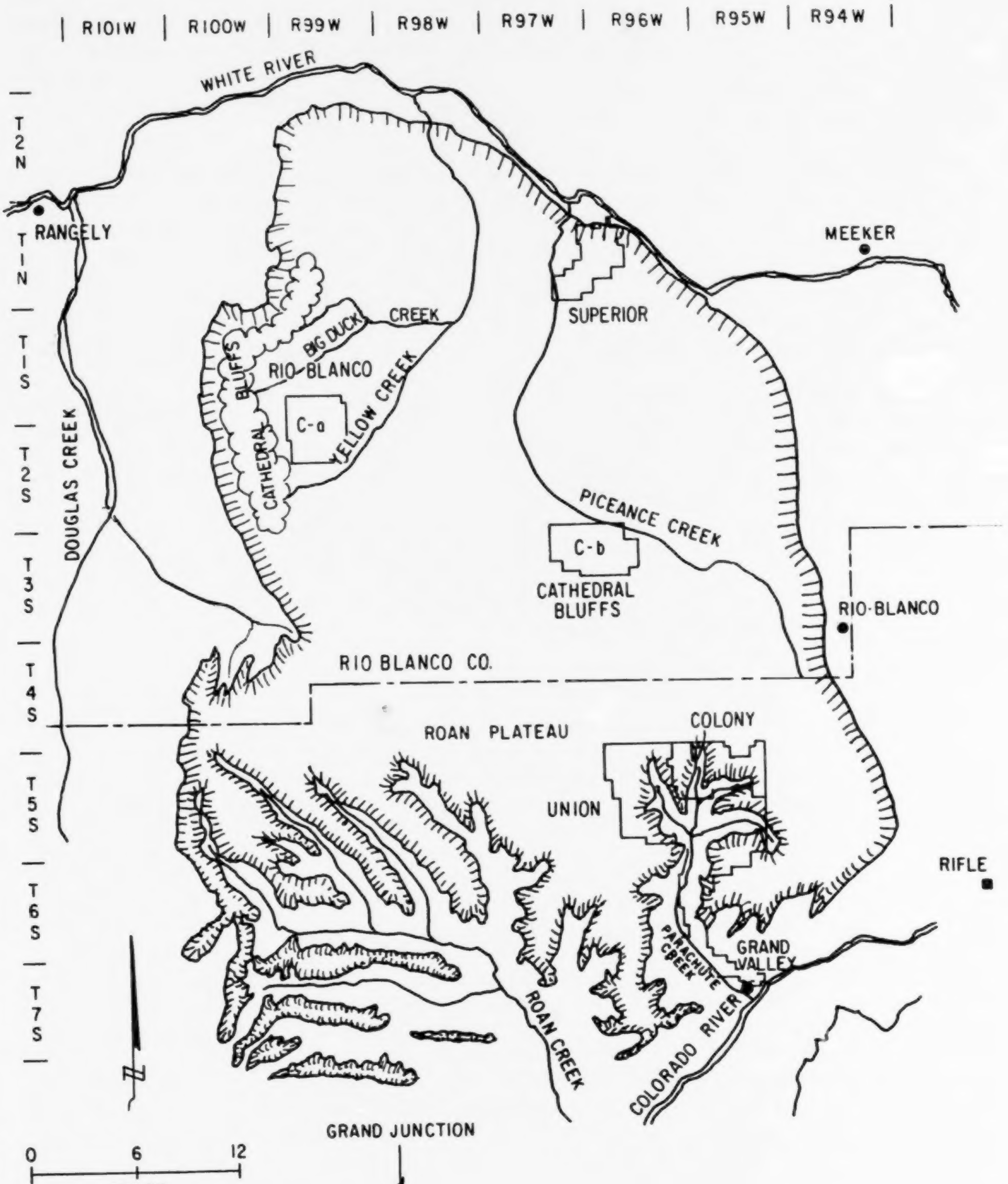
Tract C-a covers approximately 5,100 acres in Rio Blanco County, Colorado. It is located on the west flank of the Piceance Creek Basin, about 5 miles east of Cathedral Bluffs between Yellow Creek and Big Duck Creek, as illustrated in Figure 2.1-1.

Valleys and ridges crossing Tract C-a originate to the southwest in the vicinity of Cathedral Bluffs. Most of the tract is drained by Corral and Box Elder Gulches, which eventually join Stake Springs Draw to form Yellow Creek. Yellow Creek initially flows northeast, but curves to the northwest before emptying into the White River, about 20 miles north of the tract at an elevation of some 5,500 feet. The first principal drainage north of the tract is Big Duck Creek, which flows into Yellow Creek about 7 miles northeast of the tract. On the south, Ryan Gulch passes within about 2.5 miles of the tract's southern boundary before converging with Piceance Creek, about 10 miles due east of the tract where the elevation is about 6,100 feet. Tract C-a is located approximately 20 miles southeast of Rangely, 35 miles southwest of Meeker, and some 75 miles due north of Grand Junction (Gulf Oil Corp. and Standard Oil Co. [Indiana], March 1976).

Two basic weather systems affect precipitation on Tract C-a. Frontal systems generally result in widespread, uniform precipitation. Convection systems or thunderstorms result in erratic patterns of precipitation over an area of a few square miles. Annual precipitation on the tract (measured at Stake Springs Draw) for the years 1975 and 1976 measured 13.25 and 11.83 inches, respectively. Ambient temperatures are moderate during the spring, summer and fall; winter minimum temperatures are low. Gradient winds are the prevailing westerlies which occur all year, interrupted only occasionally by the passage of frontal systems. In the absence of strong gradient winds, the terrain produces local meteorological conditions (Gulf Oil Corp. and Standard Oil Co. [Indiana], May 1977).

Peak stream flows usually occur after spring snowmelt (March-April) and lows occur in late summer or early fall (August-November). Records kept from 1974 to 1976 indicated that both Corral Gulch (east) and Yellow Creek sustained baseflow, with Yellow Creek having higher discharges (averaging approximately 1,150 acre-feet annually). Corral Gulch (east) had an average annual discharge of 450 acre-feet over this period. Box Elder Gulch and Corral Gulch (west) do not sustain baseflows; however, both showed effects of snowmelt. A water analysis of Yellow Creek near the White River, conducted over an 8-month period in 1976, showed an average of 2,650 mg/l TDS and 1,475 mg/l bicarbonate. A similar analysis of the water in Corral Gulch (east) indicated a TDS content of 795 mg/l with 455 mg/l bicarbonate. Among the stream reaches on Tract C-a, iron, pH, and total dissolved solids (TDS) exceed suggested drinking water limits. Along the lower section of Corral Gulch on the tract, and in Yellow Creek, groundwater inflows cause increases in hardness, fluoride and sodium (Gulf Oil Corp. and Standard Oil Co. [Indiana], May 1977).

It is probable that some of the springs in the area supplying perennial water flow are fed by the alluvial aquifers. Along the main fork of Corral


R101W
R100W
R99W
R98W
R97W
R96W
R95W
R94W
T2N
T1N
T1S
T2S
T3S
T4S
T5S
T6S
T7S
WHITE RIVER
RANGELY
MEEKER
SUPERIOR
BIG DUCK CREEK
BLUFFS
RIO BLANCO
YELLOW CREEK
C-a
CATHEDRAL
DOUGLAS CREEK
PICEANCE CREEK
C-b
CATHEDRAL BLUFFS
RIO-BLANCO
RIO BLANCO CO.
ROAN PLATEAU
COLONY
UNION
RIFLE
PARACHUTE CREEK
GRAND VALLEY
ROAN CREEK
COLORADO RIVER
N
GRAND JUNCTION
0
6
12
MILES

SOURCE: DRI based on Gulf Oil Corp. and Standard Oil Co. (Indiana), March 1976

FIGURE 2.1-1 LOCATION OF TRACT C-a IN PICEANCE CREEK BASIN

Gulch, Box Elder Gulch, and Stake Springs Draw, the thickness of saturated alluvium ranges from 12 to 54 feet and averages 27 feet. Comparison of aquifer water level and stream flow indicates that the alluvial aquifers are very closely related to regional surface water. Springtime rises in alluvial aquifer water levels result from infiltration of snowmelt.

An extensive deep aquifer system also exists in the Piceance Creek Basin. The system mainly consists of two artesian bedrock aquifers--the upper aquifer and the lower aquifer. Although the impermeable Mahogany oil shale zone separates the two aquifers, they are interconnected in some places via natural faults and fracturing; however, there is little interconnection under Tract C-a except for the northeast corner. The aquifer thickness varies from 100 to 400 feet, with 220 feet being the average for both aquifers, and together they contain 25 million acre-feet of water.

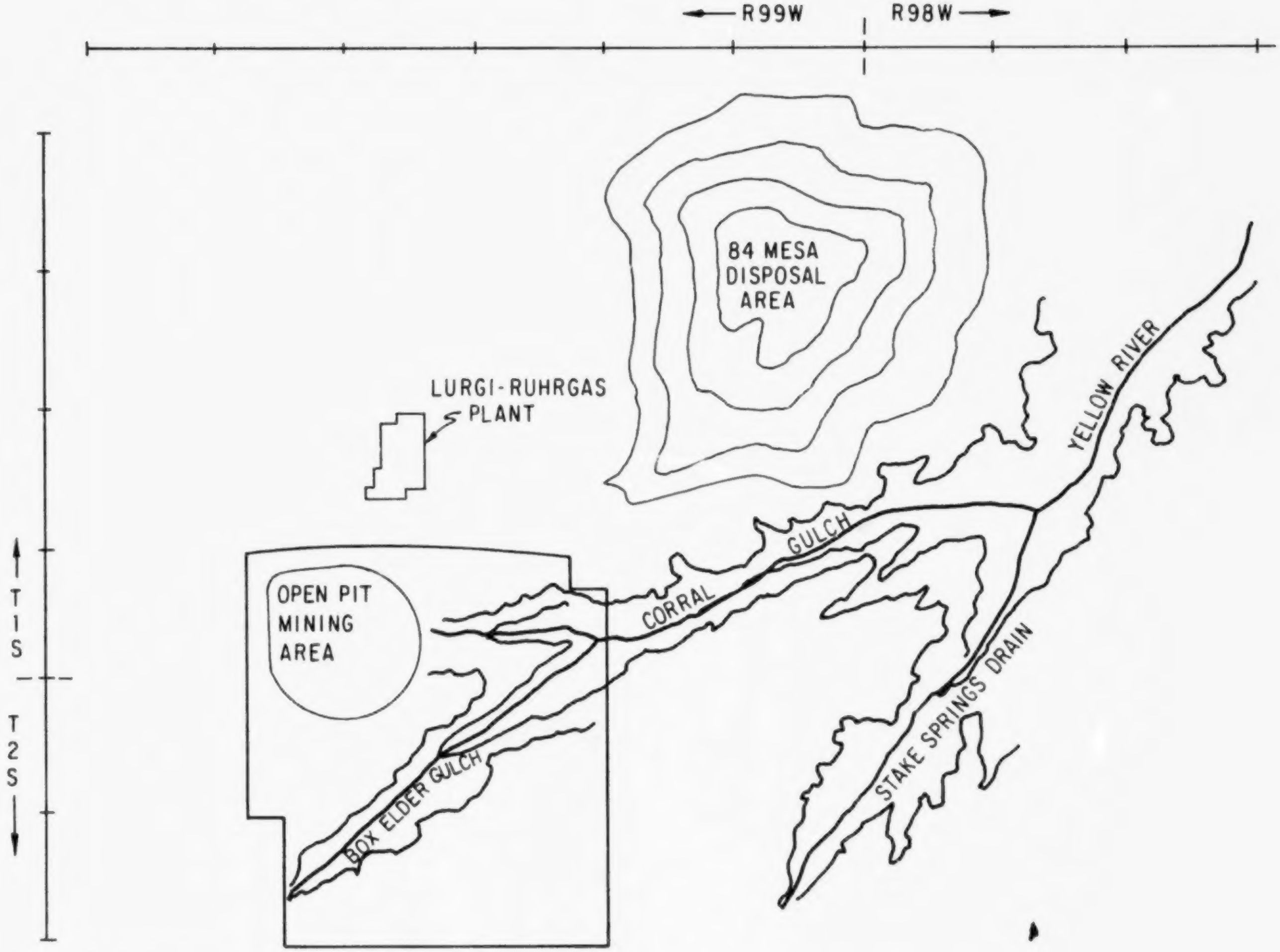
Some significant differences can be observed in the lower and upper aquifers. For example, the gradient of the lower aquifer is much flatter than the gradient of the upper aquifer. One cause of this difference may be the much higher transmissivity of the lower aquifer. Another major difference is that the upper aquifer discharges directly into Yellow and Piceance Creeks, while the lower aquifer must discharge by upward leakage to the upper aquifer. This slow, diffuse discharge over a large area should result in a region near the center of the basin over which the gradient is nearly flat. The middle of Tract C-a appears to be the border of this discharge area (Slawson, April 1980).

2.1.2 Description of the Plant Complex

Figure 2.1-2 shows the location of the off-tract disposal area, processing facilities, and open pit mine, with respect to Tract C-a. The disposal area located to the northeast of the tract is reserved for the wastes generated during the initial 30 years of tract development. The wastes produced during the 20-year project life will be placed back in the pit.

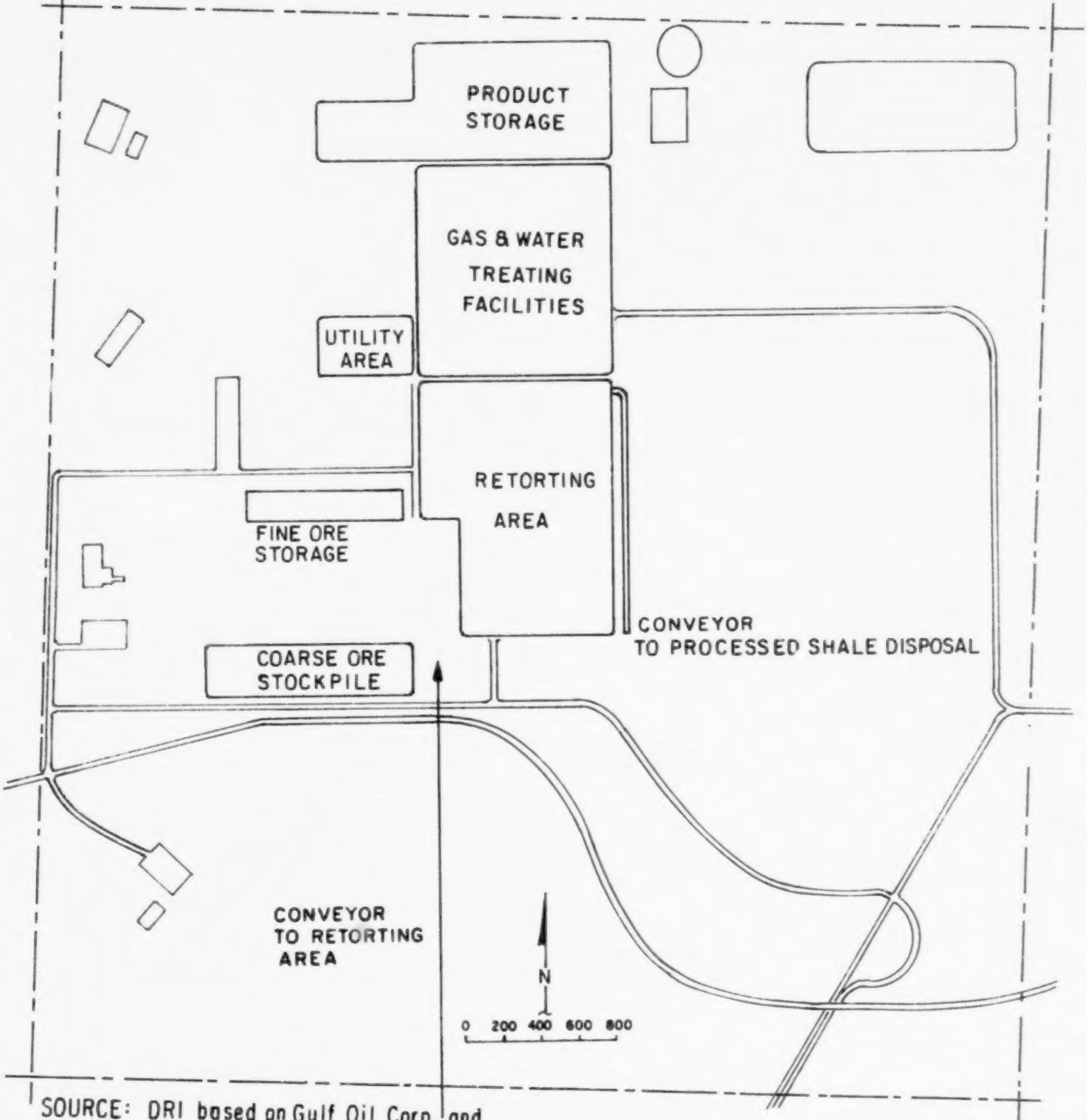
The processing facilities will be situated off of the tract, to the north. Figure 2.1-3 depicts a plot plan for the facilities, which will include 13 Lurgi processing trains with the same number of product recovery sections. The secondary and tertiary crushers will be located at the plant site; however, the primary crushers will be placed in the pit itself. Overland conveyors will transport the raw shale to the plant and carry the processed shale from the plant back to the pit. The raw shale stockpile (open) and fine shale storage bin (enclosed) will also be suitably located at the plant site. Other pertinent processing facilities on the plant plot will include gas and water (process as well as mine) treatment, product tankage and pipelines, utility area, shop and warehouse, etc.

The open pit will begin in the northwest quadrant of the tract. Figure 2.1-4 presents a detailed schematic of the pit at the beginning of the project (after the initial 30-year development). A partial geologic cross section of the pit is shown in Figure 2.1-5; as the figure illustrates, the pit will intercept the upper and lower aquifers located under the tract.


R99W
R98W
84 MESA
DISPOSAL
AREA
LURGI-RUHRGAS
PLANT
YELLOW RIVER
GULCH
CORRAL
OPEN PIT
MINING
AREA
STAKE SPRINGS DRAIN
BOX ELDER GULCH
T
1
S
T
2
S

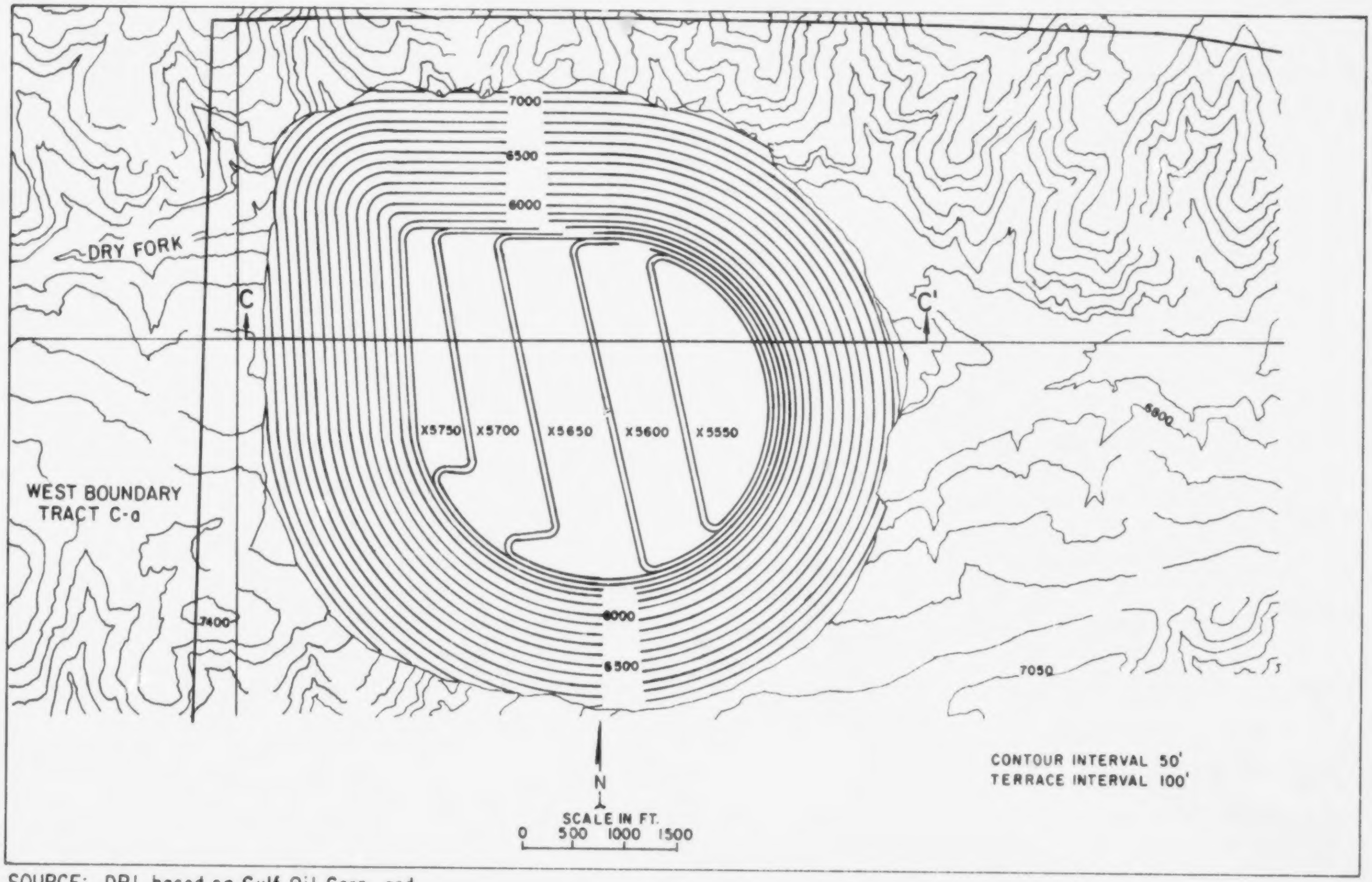
SOURCE: DRI based on Gulf Oil Corp. and
Standard Oil Co. (Indiana), March 1976

FIGURE 2.1-2 PLANT COMPLEX


PRODUCT STORAGE
GAS & WATER TREATING FACILITIES
UTILITY AREA
RETORTING AREA
FINE ORE STORAGE
COARSE ORE STOCKPILE
CONVEYOR TO PROCESSED SHALE DISPOSAL
CONVEYOR TO RETORTING AREA
N
0 200 400 600 800

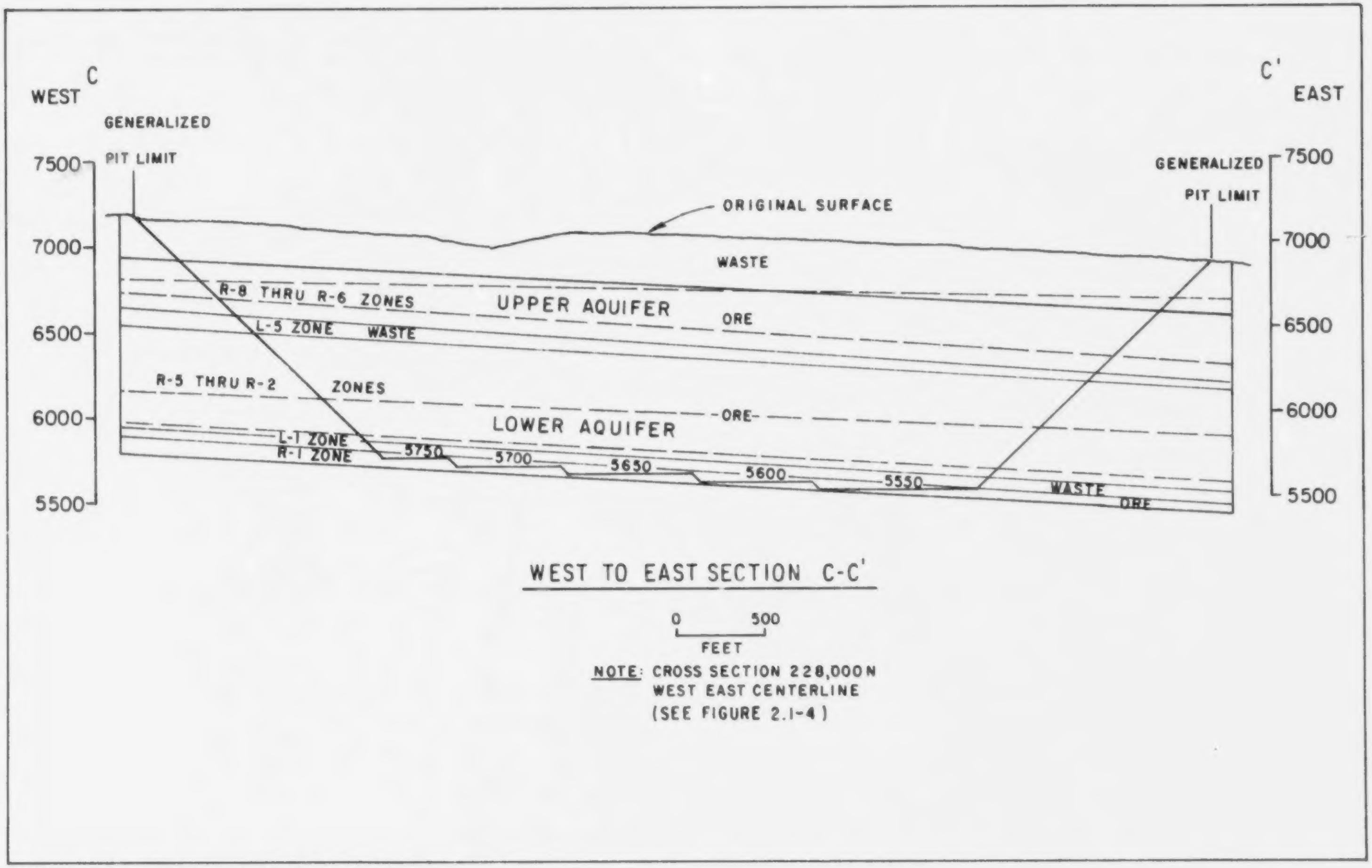
SOURCE: DRI based on Gulf Oil Corp. and Standard Oil Co. (Indiana), March 1976

FIGURE 2.1-3 PLOT PLAN FOR PROCESSING FACILITIES


7000
6500
6000
DRY FORK
C
C'
X5750
X5700
X5650
X5600
X5550
6800
WEST BOUNDARY
TRACT C-a
7400
6000
6500
7050
N
SCALE IN FT.
0 500 1000 1500
CONTOUR INTERVAL 50'
TERRACE INTERVAL 100'

SOURCE: DRI based on Gulf Oil Corp. and
Standard Oil Co. (Indiana), March 1976

FIGURE 2.1-4 30-YEAR OPEN PIT DESIGN


WEST C
C' EAST
GENERALIZED PIT LIMIT
GENERALIZED PIT LIMIT
7500
7000
6500
6000
5500
ORIGINAL SURFACE
WASTE
R-8 THRU R-6 ZONES
UPPER AQUIFER
ORE
L-5 ZONE
WASTE
R-5 THRU R-2 ZONES
ORE
LOWER AQUIFER
L-1 ZONE
R-1 ZONE
5750
5700
5650
5600
5550
WASTE
ORE
WEST TO EAST SECTION C-C'
0
500
FEET
NOTE: CROSS SECTION 228,000N
WEST EAST CENTERLINE
(SEE FIGURE 2.1-4)

SOURCE: DRI based on Gulf Oil Corp. and Standard Oil Co.(Indiana), March 1976

FIGURE 2.1-5 30-YEAR PIT CROSS SECTION

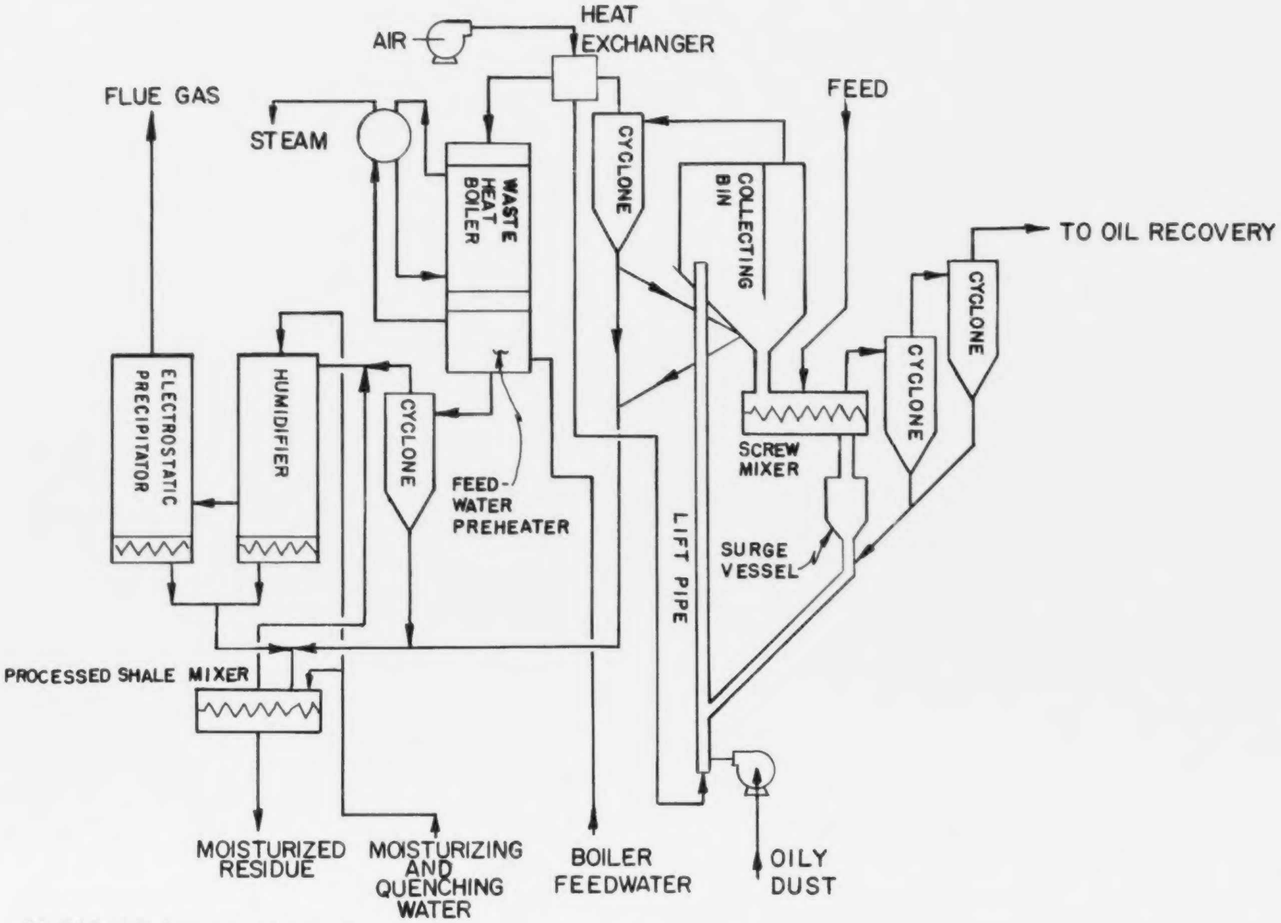
2.1.3 Description of the Retorting Process

A detailed description of the Lurgi retorting process is presented below. Unit flow diagrams describing the operation of other processing units in the integrated plant complex are presented in Section 3.

Lurgi-Ruhrgas Retorting Process--

A schematic for the Lurgi retorting process is shown in Figure 2.1-6. Initial crushing in the pit reduces the size of the run-of-mine shale to minus 8 inches. Secondary and tertiary crushing further reduce the shale size to minus 1/4 to 1/3 inches. The crushed oil shale is fed through a feed hopper to a double screw mixer, where four to eight times its weight of a hot (1,200-1,300°F) circulating heat carrier, such as sand or processed shale from the collecting bin, is thoroughly mixed in, thus heating the entire mixture to approximately 950-1,000°F within a few seconds. At this temperature, pyrolysis of the kerogen in the oil shale occurs, resulting in the production of retort gas, shale oil vapor and water vapor. The circulating heat carrier and the partially retorted shale are then dropped from the screw mixer into the surge vessel, where residual oil components are distilled off. The mixture of heat carrier and retorted shale residue is passed to the lower section of the lift pipe, where combustion air (preheated to 450-900°F) is introduced, raising the mixture pneumatically to the collecting bin (TRW and DRI, 1975-1978; York, June 13, 1980). Essentially all available organic carbon contained in the retorted shale residue is burned in the lift pipe. Supplemental fuel may be added to the bottom of the lift pipe to sustain the combustion of the organic residue when processing leaner oil shales. Also, at the high lift pipe temperature, a moderate amount of carbonate decomposition occurs in the processed shale. At the top of the lift pipe, the hot, burned shale is separated from the flue gases in the collecting bin. Fines are carried out of the collecting bin with the flue gas stream. The coarse-grained shale residue accumulates in the lower section of the collecting bin and flows continuously to the mixer. Partial removal of the solids to prevent accumulation in the collecting bin may be required if the fines carry-over is not sufficient. If the shale disintegrates to the extent that more fines are produced than expected, an additional coarse-grained heat carrier, such as sand or low-grade shale, may be needed. The combustion air supplied to the lift pipe is preheated by counter-current heat exchange with the flue gas stream in the preheat section of the waste heat boiler. The calcined minerals in the burned shale combine with the sulfur dioxide produced by combustion of the organic sulfur to form calcium and magnesium sulfites and sulfates (Rio Blanco Oil Shale Co., February 1981).

The pyrolysis products stream containing shale fines is withdrawn at the end of the screw mixer and passed through two series-connected cyclones to a product recovery section. The fines are separated in these cyclones and returned to the recycle system. The vapor stream then passes through a sequence of three scrubbing coolers (not shown; see Figure 3.3-4 in Section 3). The first scrubbing cooler is designed to operate at a high temperature (~350°F) and to remove dust from the gas stream by condensation of heavier oil fractions. Circulation of the condensed heavy oil through the


AIR
HEAT EXCHANGER
FLUE GAS
STEAM
FEED
WASTE HEAT BOILER
CYCLONE
COLLECTING BIN
TO OIL RECOVERY
CYCLONE
CYCLONE
ELECTROSTATIC PRECIPITATOR
HUMIDIFIER
CYCLONE
SCREW MIXER
FEED-WATER PREHEATER
LIFT PIPE
SURGE VESSEL
PROCESSED SHALE MIXER
MOISTURIZED RESIDUE
MOISTURIZING AND QUENCHING WATER
BOILER FEEDWATER
OILY DUST

SOURCE: DRI based on Schmalfeld,
July 1975

FIGURE 2.1-6 LURGI- RUHRGAS OIL SHALE RETORTING PROCESS

scrubber aids in removal of the dust. A dusty heavy oil is obtained at this point. The operating temperature of this scrubber is controlled by introducing and evaporating the gas liquor through the scrubber. The amount of heavy oil and its properties can be varied in this fashion. In the next scrubbing cooler, further condensation of the oil takes place at a temperature above the dew point of water to produce a water-free middle oil (Schmalfeld, July 1975). Final cooling of the gas produces an aqueous gas condensate and a light oil fraction. The light oil is separated from the condensate or gas liquor in an oil/water separator. Partial amounts of ammonia and sulfur dioxide in the gas stream are absorbed in the gas liquor. If necessary, further removal of these species from the gas can be achieved by circulating more of the gas condensate through the third scrubber. Finally, the gas is scrubbed with a lean oil in the naphtha scrubber to recover naphtha and noncondensable gases, as deemed desirable. Residual H_2S may be removed from the remaining gas by one of several methods. The gas liquor may also be cleaned before reuse or discharge.

The flue gas stream in the lift pipe is dedusted in a cyclone after leaving the collecting bin; it is then routed through a heat exchanger for preheating of combustion air, a waste heat boiler to produce process steam, another cyclone, and a humidifier or flue gas conditioner. The flue gas stream is cooled somewhat and conditioned in the humidifier by adding steam generated during processed shale quenching. After humidification and cooling, residual dust is removed from the flue gas stream using an electrostatic precipitator and discharged into a processed shale quencher/moisturizer where more water is added to cool the solids. The processed shale residue, cooled to ~200°F, is moisturized to a suitable moisture content and discarded.

The dusty heavy oil obtained from the first scrubbing cooler is thinned with an available lighter oil from the process and subjected to centrifugation to remove the dust. The clean oil is stabilized by vaporizing the light oil components and then sent for storage. The recovered light oil is recycled to the process and the dust is fed to the bottom of the lift pipe and burned.

2.2 POLLUTION CONTROL CASE STUDY

The case study examined in this manual is based primarily on information published by the developers (Gulf Oil Corp. and Standard Oil Co. [Indiana], March 1976; Rio Blanco Oil Shale Co., February 1981). The pollution control approaches analyzed for the commercial plant activities, such as mining and crushing, Lurgi flue gas discharge, retort gas treatment, gas liquor treatment, excess mine water discharge, and solid waste disposal, are intended to serve as illustrative examples only and should not limit consideration of other alternatives.

Since standard industry practices are adopted for various minor treatments, these technologies are not discussed in detail (e.g., boiler feedwater makeup treatment). The impact on the cost of treatment as a result of variations in the pollution control strategy in other processing areas is assessed, but a detailed analysis of the treatment technology itself is not presented.

2.2.1 Key Features of Pollution Control

The primary feature of this manual is that it provides an opportunity to analyze the impacts resulting from open pit mining. This mining technique was selected by the Tract C-a developer because the oil shale deposits are technologically and economically amenable to this type of recovery and, furthermore, it affords nearly complete mining of the resource. As a disadvantage, however, large amounts of overburden and subgrade ore also must be removed along with the retortable oil shale, making it the largest mining operation of its kind in the world.

All of the mined materials are crushed in the pit and transported for disposal or processing. The transportation is achieved by an extensive network of overland enclosed conveyors equipped with dust control systems such as baghouses and water and foam sprays. Further crushing of the oil shale takes place at the plant located off site, to the north of the tract. Because of the size and extent of this materials handling system, the plant uses an unusually high number of baghouses and dust suppression devices.

The commercial open pit will intercept two extensive deep aquifer systems which lie under the tract. These aquifers slope gently to the northeast toward the center of the hydrologic basin of Piceance Creek; the waters are mostly stagnant, as the aquifer recharge occurs primarily from precipitation along the basin margins, and discharge is by release to Piceance Creek. The effect of intersecting the aquifers in the pit will be the tendency of the waters to flow from all directions into the pit. Thus, the aquifers would need to be dewatered throughout the project life to avoid infiltration of water into the pit. The transmissivity, storage coefficient, and thickness data for the two aquifers suggest a dewatering rate of approximately 16,500 gpm (Gulf Oil Corp. and Standard Oil Co. [Indiana], March 1976). About 70 dewatering wells around the periphery of the pit may be required for dewatering (these wells are assumed to be in place before the process analysis in this case study begins). Although the process will have zero water discharge in terms of process waters, the result of dewatering is that an excess of mine water will remain after the process needs have been satisfied. Since this will necessitate disposal of the excess mine water, the overall plant will no longer be a zero discharge facility.

The plant complex considered here will not burn any fuels for power or steam generation. Electricity will be obtained from outside sources, and sufficient steam will be generated by the Lurgi process. Thus, there will not be any major flue gas sources besides the stack in the processed shale discharge system. Since the retort gas is prepared for selling purposes, its cleanup does not qualify as pollution control. If the gas were being cleaned for on-site use, the cleaning process would have qualified as a pollution control measure. Nonetheless, any treatment of the tail streams before discharge into the environment is considered as pollution control.

Another unusual feature of this study is that it assumes full-scale development of the open pit, with this operation occurring during the 30 years prior to the starting point of this analysis. In such a case, the current plant will be the second oil shale processing facility and many of the pollution controls and other equipment, such as baghouses, dust suppression system, water management system, storage tanks, etc., will already be in place. For the economic analysis, however, this study assumes that all pollution control measures are newly installed during the start-up period for the second, or current, plant. Furthermore, the wastes generated during the initial 30 years have already been disposed of on a remote location, and the environmental and cost issues associated with the off-site disposal are not addressed.

2.2.2 Pollution Control Case Study

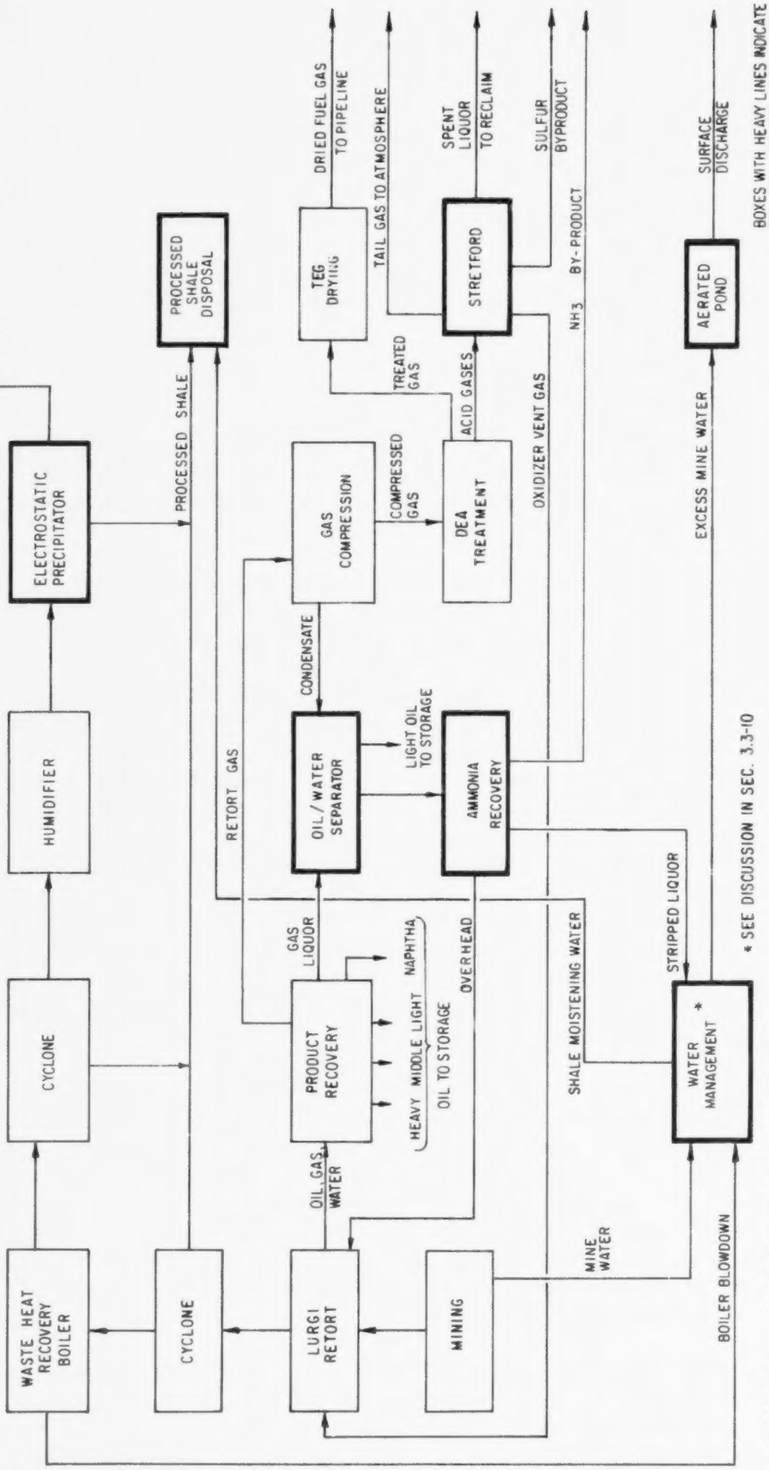
A block flow diagram for the basic processing and pollution control system for the case study analyzed in this manual is presented in Figure 2.2-1. The pollution control areas are highlighted in the diagram by heavy lines. A brief overview of the entire process follows.

Mining of the oil shale is performed by the open pit method. The fugitive dust generated during this operation is controlled with water and foam sprays. The run-of-mine oil shale, subore, and overburden are crushed to a size of 6 to 8 inches in individual crushers located in the pit itself. The crushing operation also generates particulates which are controlled by baghouses installed on the crushers. The crushed subore and overburden are sent for disposal in the back of the pit, while the oil shale is transported to the surface using enclosed conveyors. These conveyors are equipped with baghouses and dust suppression devices to control the particulates generated at transfer points. The diesel-powered machinery used in mining and disposal activities is equipped with catalytic converters to control the carbon monoxide and unburned hydrocarbons in the exhaust gases.

The primary crushed shale is further reduced in size by secondary and tertiary crushing and then fed to the retorts. The crushing, screening, transporting, storage, and feeding operations generate airborne particulates which are also controlled by baghouses.

In the Lurgi retorts, the raw shale is pyrolyzed by mixing it with a portion of previously processed hot shale. The vaporized oil and gas products from the pyrolysis are sent to the oil and gas recovery section of the plant, while the retorted shale is sent to the lift pipes where the residual organic matter on the retorted shale is incinerated to generate the heat necessary for retorting the raw shale. A flue gas is also produced as a result of incineration. As mentioned previously, a portion of the processed shale is recycled to the retorts, while the remainder is passed along with the flue gas through the discharge system.

In the discharge system, the flue gas and entrained processed shale particles are separated from each other via a series of cyclones, waste heat recovery system, humidifier, etc. The flue gas is then passed through an electrostatic precipitator to remove the residual particulates and is


WASTE HEAT RECOVERY BOILER
CYCLONE
CYCLONE
HUMIDIFIER
ELECTROSTATIC PRECIPITATOR
FLUE GAS
PROCESSED SHALE
PROCESSED SHALE DISPOSAL
LURGI RETORT
OIL, GAS, WATER
PRODUCT RECOVERY
HEAVY MIDDLE LIGHT NAPHTHA
OIL TO STORAGE
GAS LIQUOR
RETORT GAS
OIL / WATER SEPARATOR
LIGHT OIL TO STORAGE
CONDENSATE
GAS COMPRESSION
COMPRESSED GAS
DEA TREATMENT
TREATED GAS
ACID GASES
TEG DRYING
DRIED FUEL GAS TO PIPELINE
TAIL GAS TO ATMOSPHERE
STRETFORD
SPENT LIQUOR TO RECLAIM
SULFUR BYPRODUCT
OXIDIZER VENT GAS
NH3 BY-PRODUCT
MINING
MINE WATER
OVERHEAD
AMMONIA RECOVERY
SHALE MOISTENING WATER
STRIPPED LIQUOR
WATER MANAGEMENT *
BOILER BLOWDOWN
EXCESS MINE WATER
AERATED POND
SURFACE DISCHARGE

BOXES WITH HEAVY LINES INDICATE POLLUTION CONTROL ACTIVITIES

* SEE DISCUSSION IN SEC. 3.3-10

SOURCE: DRI based on Gulf Oil Corp. and Standard Oil Co. (Indiana) March 1976, and Rio Blanco Oil Shale Co., February 1981

FIGURE 2.2-1 PROCESS FLOW DIAGRAM

eventually vented to the atmosphere. The processed shale separated along these steps is sent to the processed shale mixer for quenching and proper moisturizing before final disposal.

In the oil and gas recovery unit, the products of pyrolysis are separated into heavy, middle and light oils, naphtha, gas liquor and noncondensable retort gas. The oils and naphtha are sent for storage, while the gas liquor and retort gas are treated further. Fugitive hydrocarbons emanate from the product storage and these are controlled primarily by employing floating roof storage tanks.

The retort gas is cleaned for the purpose of selling it. The gas is first compressed to remove much of the moisture and ammonia, then subjected to treatment by diethanolamine and triethylene glycol, which remove the acidic components and the residual moisture, respectively, from the compressed gas. The clean, dry gas is then sent to the pipeline. Since these are processing steps, they are not considered as pollution control measures.

The acid gas obtained from the diethanolamine regeneration is treated by the Stretford process, which converts the H_2S in the gas to elemental sulfur. The clean gas is then vented to the atmosphere. Since a direct release of the acid gas (before treatment by the Stretford process) would create pollution, the acid gas treatment is considered a pollution control measure.

The gas liquor from the oil and gas recovery section is subjected to oil/water separation, but it still contains dissolved ammonia and sulfur compounds and its direct discharge or use may also create pollution. Therefore, the ammonia and dissolved volatile compounds from the liquor are removed by an ammonia recovery process. The treated water is then used for processed shale moisturizing.

Dewatering of the two aquifers under Tract C-a is necessary in order to keep the pit dry. The water thus obtained is used to satisfy the processing requirements, but the total amount of water is in excess of that needed. As a result, the excess water must be discarded. The excess water is retained in an aeration pond for a period of one week to oxidize the organic matter; it is then discharged into natural surface water streams.

The wastes generated from mining and processing of the oil shale are backfilled into the pit. During the active backfilling operation, the fugitive dust generated due to the vehicular traffic is controlled by the application of water and foam sprays. After establishing the final contours, the surface of the backfilled wastes is covered with topsoil and revegetated.

Overall water management activities consist of satisfying the process steam and cooling water needs, as well as efficient management of the aqueous waste effluents. Properly treated mine water is used as the boiler feedwater to produce the steam in the Lurgi waste heat recovery boiler. Treated mine water is also used as cooling water, process makeup water, cooling tower makeup water, etc. Minor wastes generated from the water treatments are equalized in a holding pond and then used for processed shale moisturizing.

The mine water is also used for dust control, revegetation, sanitary uses, and as service and fire water.

Proper maintenance practices are used to reduce the fugitive hydrocarbons emanating from valves, pumps, etc.

2.3 SUMMARY OF POLLUTION CONTROL TECHNOLOGIES AND COST

The control technologies examined in the case study are summarized in Table 2.3-1. As a means of organizing the presentation, the plant complex is divided into different areas of processing activities and pollution control. It should be noted that the control technologies examined here are not the only choices available, nor are they necessarily sufficient for pollution control; rather, they are merely examples from broad classes of technologies. These examples have been examined on the basis that they have been proposed at one time or another by oil shale developers. Additionally, good vendor guarantees and cost data on these technologies were available for the economic analysis.

Throughout this analysis of the Lurgi-Open Pit project, the distinction between process and pollution control is not always clear. For example, the diethanolamine treatment of the retort gas could be considered a pollution control measure because it affords removal of H_2S. However, since the main purpose behind the treatment is to sell the gas and not to use it on site, the treatment is considered a processing step. Similarly, boiler feedwater treatment, cooling water treatment, source water clarification, etc., are listed as pollution control measures, when they may also be classified as process related activities. In some such instances--for example, the cooling water treatment--only the cost increase due to the pollution control activities is included, but this distinction is not always possible. Consideration of an activity as a pollution control or as a process related activity becomes important when calculating the total cost of pollution control. Because all of the borderline activities are classified as pollution control, the user of this manual should be made aware that the total pollution control costs are conservatively stated due to the inclusion of activities which could also be considered process related.

Table 2.3-2 lists the control technologies examined in the case study, along with information describing location, control function and size. More detailed design information for the technologies is presented in Section 5. A discussion of other possible control choices is also given in that section.

Table 2.3-3 summarizes the costs of air pollution control and water management and pollution control for the case study analyzed for the Lurgi-Open Pit facility. The costs for solid waste management are not included in the table because of insufficient information regarding the developer's plans for solid waste disposal. Detailed engineering costs for the technologies analyzed and the cost computation methodology are presented in Section 6.

TABLE 2.3-1. SUMMARY OF POLLUTION CONTROL TECHNOLOGIES

Areas of Control						
Mining, Crushing and Materials Handling	Retorting Treatment	Retort Gas Treatment	Mine Water Treatment	Gas Liquor Treatment	Steam, Power Generation	Solid Waste Management
Baghouses Water and Foam Sprays	Electrostatic Precipitator for Flue Gas	Stretford for the Acid Gases from DEA Unit	Aerated Pond	Ammonia Recovery Process	Steam Generation inherent to the retorting process; no control necessary	Open Pit Backfilling

Source: DRI based on information from Gulf Oil Corp. and Standard Oil Co. (Indiana), March 1976, and Rio Blanco Oil Shale Co., February 1981.

TABLE 2.3-2. INVENTORY OF MAJOR POLLUTION CONTROL TECHNOLOGIES

Type of Control (Number of Units)	Origin of Material Controlled	Material Controlled	Flow Rate Each Unit	Processing Activity Area
Water and Foam Sprays	Mine, Open Stockpiles	Raw and Processed Shale Dust	--	Mining, etc.
Fabric Filter (2)	Primary Crushers (ore)	Raw Shale Dust	61,100 ACFM	Mining, etc.
Fabric Filter (1)	Primary Crushers (subore)	Subore Shale Dust	12,200 ACFM	Mining, etc.
Fabric Filter (1)	Primary Crushers (overburden)	Overburden Dust	63,800 ACFM	Mining, etc.
Fabric Filter (2)	Conveyor to Stockpile	Raw Shale Dust	36,300 ACFM	Mining, etc.
Fabric Filter (3)	Raw Shale Conveyor Transfer Points	Raw Shale Dust	40,500 ACFM	Mining, etc.
Fabric Filter (2)	Conveyor to Secondary Crushers	Raw Shale Dust	20,200 ACFM	Mining, etc.
Fabric Filter (8)	Secondary Crushers	Raw Shale Dust	69,800 ACFM	Mining, etc.
Fabric Filter (2)	Conveyor to Secondary Screens	Raw Shale Dust	20,200 ACFM	Mining, etc.
Fabric Filter (8)	Secondary Screens	Raw Shale Dust	69,800 ACFM	Mining, etc.
Fabric Filter (2)	Conveyor to Tertiary Crushers	Raw Shale Dust	20,200 ACFM	Mining, etc.
Fabric Filter (9)	Tertiary Crushers	Raw Shale Dust	69,800 ACFM	Mining, etc.
Fabric Filter (4)	Conveyor to Tertiary Screens	Raw Shale Dust	20,200 ACFM	Mining, etc.
Fabric Filter (9)	Tertiary Screens, Both Sets	Raw Shale Dust	69,800 ACFM	Mining, etc.
Fabric Filter (2)	Conveyor to Fine Ore Storage	Raw Shale Dust	20,200 ACFM	Mining, etc.
Fabric Filter (1)	Fine Ore Storage	Raw Shale Dust	28,800 ACFM	Mining, etc.
Fabric Filter (2)	Conveyor to Retort Feed Hoppers	Raw Shale Dust	20,200 ACFM	Mining, etc.
Fabric Filter (4)	Retort Feed Hoppers	Raw Shale Dust	53,100 ACFM	Mining, etc.
Fabric Filter (13)	Conveyor to Retorts	Raw Shale Dust	18,400 ACFM	Mining, etc.
Electrostatic Precipitator (13)	Flue Gas Discharge System	Processed Shale Dust	293,700 ACFM	Processed Shale Removal (pyrolysis)
Fabric Filter (2)	Processed Shale Conveyor Transfer Points	Processed Shale Dust	32,300 ACFM	Processed Shale Disposal
Fabric Filter (3)	Processed Shale Load-out Hoppers	Processed Shale Dust	21,500 ACFM	Processed Shale Disposal

(Continued)

TABLE 2.3-2 (cont.)

Type of Control (Number of Units)	Origin of Material Controlled	Material Controlled	Flow Rate Each Unit	Processing Activity Area
Stretford (1)	DEA Unit	H_2S	10,500 ACFM	Retort Gas Treatment
Oil/Water Separator (1)	Lurgi Product Recovery	Oil Emulsion in Water	586 gpm	Gas Liquor Treatment
Ammonia Recovery (1)	Oil/Water Separator	H_2S, NH_3 and Volatile Organics in Water	586 gpm	Gas Liquor Treatment
Mine Water Clarifier (1)*	Underground Aquifers	Suspended Matter	16,500 gpm	Mine Water Treatment
Aeration Pond (1)	Mine Water Clarifier	Dissolved Organics	8,330 gpm	Excess Mine Water Treatment
Floating Roof Storage Tanks (2)	Lurgi Product Storage	Hydrocarbons	63,140 BPSD	Miscellaneous Air Treatment
Ammonia Storage Tank (1)	Ammonia Recovery Unit	Ammonia	22.6 TPSD	Miscellaneous Air Treatment
Catalytic Converters	Diesel Equipment	CO, HC	--	Miscellaneous Air Treatment
Proper Maintenance of Valves, Pumps, etc.	Valves, Pumps, etc.	Hydrocarbons	--	Miscellaneous Air Treatment
Boiler Feedwater Treatment*	Mine Water	Dissolved Solids	43 gpm	Miscellaneous Water Treatment
Cooling Water Treatment*	Mine Water	Dissolved Solids	2,676 gpm	Miscellaneous Water Treatment
Equalization Pond	Water Treatments	Blowdowns, Runoff, Concentrates, etc.	2,525 gpm	Miscellaneous Water Treatment
Oil/Water Separator	Plant Site Storm Sewer	Plant Runoff	169 gpm	Miscellaneous Water Treatment
Runoff Collection Sumps	Waste Landfill	Leachable Compounds	--	Surface Hydrology
Runoff Collection Pumps	Runoff Collection Sumps	Leached Compounds	--	Surface Hydrology
Dust Suppression	Waste Landfill	Particulates	--	Surface Stabilization
Grubbing, Stripping, and Clearing	Waste Landfill	Soil (erosion)	--	Surface Stabilization
Reclamation and Revegetation	Waste Landfill	Soil (erosion)	--	Surface Stabilization

* The technologies marked with an asterisk (*) could be considered part of the process as well as pollution control.

Source: DRI based on information from Gulf Oil Corp. and Standard Oil Co. (Indiana), March 1976, and Rio Blanco Oil Shale Co., February 1981.

TABLE 2.3-3. POLLUTION CONTROL COST SUMMARY[a]

Control Medium	Fixed Capital Cost ($000's)	Total Annual Capital Charge ($000's)	Total Annual Operating Cost ($000's)	Total Annual Control Cost ($000's)	Per-barrel Control Cost (cents)
Air Pollution	91,042	14,747	9,013	23,760	115
Water Management and Pollution Control	7,122	1,412	2,446	3,858	19
Solid Waste Management[b]	N.D.	N.D.	N.D.	N.D.	N.D.

[a] See Section 6 for details.

[b] The solid waste management costs have not been determined (N.D.) in an integrated fashion. See Section 6 for details on individual solid waste management items.

Source: DRI.

SECTION 3

PROCESS FLOW DIAGRAMS AND FLOW RATES

Flow diagrams illustrating all operations in the Lurgi-Open Pit plant complex are presented in this section. The integrated designs shown are consistent with proposed development plans.

3.1 STRUCTURE OF THE DIAGRAMS

In order to understand the interactions throughout the plant complex, an overall flow diagram is presented first, followed by flow diagrams for each unit process. Flow rates for all major process and waste streams are indicated on each of the more detailed diagrams; flow rates for streams of an auxiliary nature, such as cooling water and steam, are included only when relevant to pollution control activities. The following symbols are used to indicate the physical state of each stream:

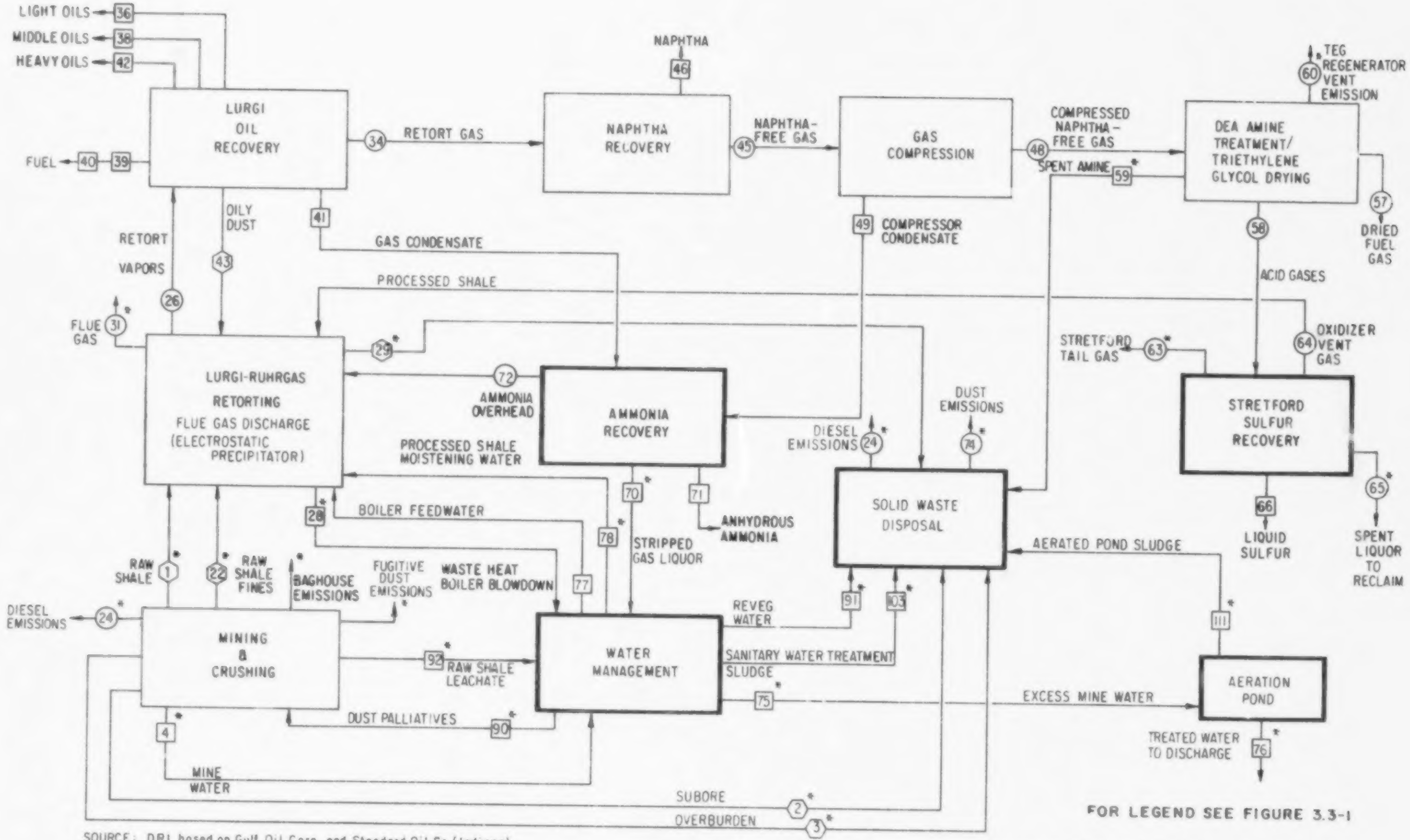
- Gases--Circles
- Liquids--Squares
- Solids--Hexagons.

A unique stream number is placed within each symbol. In addition, an asterisk (*) is placed next to the symbol for a stream if it comes into contact with the environment at any point in the process. The stream numbering system established in this section is used throughout this manual.

3.2 OVERALL PLANT COMPLEX

A flow diagram of the complete plant complex, emphasizing the waste streams produced, is presented in Figure 3.2-1. Production-scale mining of the oil shale will be accomplished utilizing the concept of a migrating open pit. Initial excavation will have begun in the northwest quadrant of Tract C-a (see Figure 2.1-2) and continued for 30 years. The waste material produced during these years will have been removed to an off-site disposal area. After the 30-year development, the pit would be sufficiently large to accommodate the simultaneous waste backfilling and mining operations. The Lurgi-Open Pit case study examines activities that occur after simultaneous backfilling and mining operations commence.

The pit boundaries at the end of the 20-year project life (following the initial 30 years) will extend south across Corral Gulch and east to near the confluence of Dry Fork and Corral Gulch. The working pit will have a depth of approximately 1,350 feet and a diameter of 7,900 feet at the surface. The


LIGHT OILS
36
MIDDLE OILS
38
HEAVY OILS
42
FUEL
40
39
LURGI OIL RECOVERY
34
RETORT GAS
NAPHTHA
46
NAPHTHA RECOVERY
45
NAPHTHA-FREE GAS
GAS COMPRESSION
48
COMPRESSED NAPHTHA-FREE GAS
SPENT AMINE
59
TEG REGENERATOR VENT EMISSION
60
DEA AMINE TREATMENT/ TRIETHYLENE GLYCOL DRYING
57
DRIED FUEL GAS
58
ACID GASES
49
COMPRESSOR CONDENSATE
RETORT VAPORS
26
OILY DUST
43
41
GAS CONDENSATE
PROCESSED SHALE
FLUE GAS
31
29
LURGI-RUHRGAS RETORTING FLUE GAS DISCHARGE (ELECTROSTATIC PRECIPITATOR)
72
AMMONIA OVERHEAD
AMMONIA RECOVERY
PROCESSED SHALE MOISTENING WATER
70
71
ANHYDROUS AMMONIA
STRIPPED GAS LIQUOR
78
BOILER FEEDWATER
28
STRETFORD TAIL GAS
63
64
OXIDIZER VENT GAS
STRETFORD SULFUR RECOVERY
66
LIQUID SULFUR
65
SPENT LIQUOR TO RECLAIM
DIESEL EMISSIONS
24
DUST EMISSIONS
74
SOLID WASTE DISPOSAL
AERATED POND SLUDGE
111
RAW SHALE
1
22
RAW SHALE FINES
BAGHOUSE EMISSIONS
FUGITIVE DUST EMISSIONS
WASTE HEAT BOILER BLOWDOWN
77
DIESEL EMISSIONS
24
MINING & CRUSHING
92
RAW SHALE LEACHATE
WATER MANAGEMENT
REVEG WATER
91
103
SANITARY WATER TREATMENT SLUDGE
75
EXCESS MINE WATER
AERATION POND
TREATED WATER TO DISCHARGE
76
DUST PALLIATIVES
90
4
MINE WATER
SUBORE
2
OVERBURDEN
3
*

FOR LEGEND SEE FIGURE 3.3-1

BOXES WITH HEAVY LINES INDICATE POLLUTION CONTROL ACTIVITIES

SOURCE: DRI based on Gulf Oil Corp. and Standard Oil Co. (Indiana), March 1976, and Rio Blanco Oil Shale Co., February 1981

FIGURE 3.2-1 PROCESS OPERATIONS AND WASTE STREAMS

slope of the pit wall will be 45°, which will be sufficient to avoid subsidence. The benches will be placed 50 feet apart.

The mined raw shale is first crushed in movable primary crushers located in the pit and then transferred to the surface by covered conveyors. Proper retort feed is then prepared by secondary and tertiary crushing and screening. The crushed shale is fed to the Lurgi retorts where it is mixed with hot, burned, processed shale, raising it to a sufficent temperature (950-1,000°F) to release a mixture of oil and high-Btu gas which moves to a recovery section of the plant. After retorting, the processed shale contains a carbon residue which is burned in a lift pipe, thereby further raising the temperature of the processed shale (1,200-1,300°F) before it is mixed with the incoming raw shale. Part of the burned, processed shale that is not recycled exits with the flue gas and is separated, quenched with water, and moistened to 19% water content before disposal.

In the Lurgi oil recovery section of the plant, three oil fractions and a high-Btu gas are recovered. A gas liquor, or condensate, with dissolved ammonia and organics is also recovered in the product recovery section.

After the stripping of naphtha, the retort gas is compressed and sent to the diethanolamine (DEA) scrubbers for removal of acid gases. The clean gas is subsequently dried in the triethylene glycol (TEG) dehydration system and sent to the pipeline. The acid gas overhead from the amine absorption system is treated in a Stretford unit to produce a sulfur by-product and a tail gas which is released to the atmosphere.

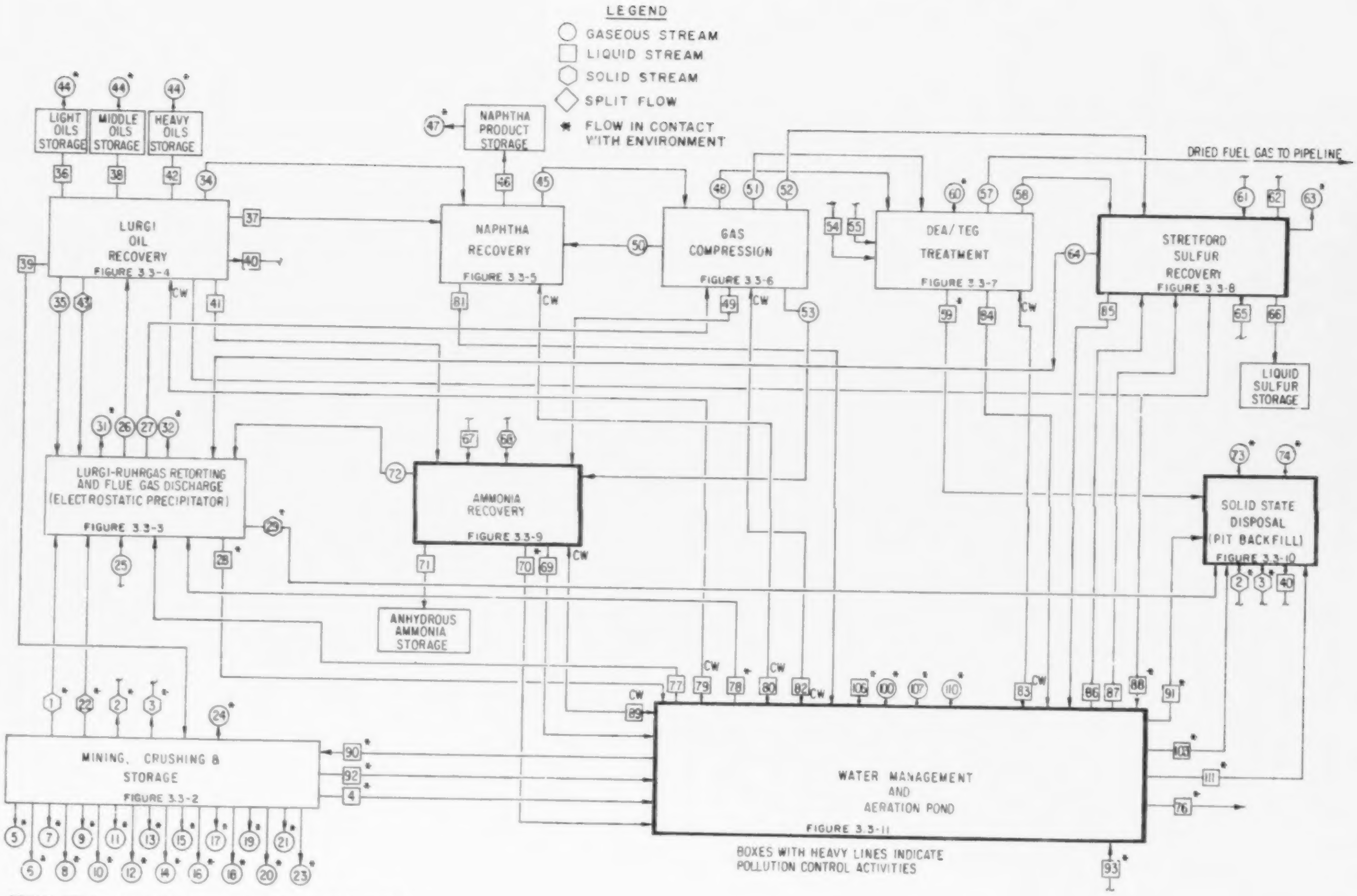
The gas liquor is sent to the ammonia recovery unit to produce anhydrous ammonia and then the stripped liquor is used for processed shale moistening.

Since the pit will intercept two underground water bodies, the upper and lower aquifers, dewatering of the mine will be necessary during the active life of the project. The amount of mine water obtained by dewatering, however, will be in excess of that needed in the plant and this excess water will need to be discharged. Aeration of the excess water to oxidize the organic material and to settle out oxidizable inorganic compounds will be carried out prior to the discharge.

The overall processing steps outlined above comprise the case study examined in this manual.

3.3 UNIT PROCESS FLOW DIAGRAMS

This section describes the operation of the Lurgi-Open Pit plant complex in more detail using flow diagrams for each unit process in the plant. Figure 3.3-1 is intended to be used as a road map showing the relationship between the unit process flow diagrams. Each box (except the product storage boxes) in the figure represents an individual flow diagram, and the appropriate figure number for each diagram is indicated. All streams are numbered as well. A complete list of all the streams, in numerical order, is included in Section 1.7, Table 1.7-1.


LEGEND
GASEOUS STREAM
LIQUID STREAM
SOLID STREAM
SPLIT FLOW
FLOW IN CONTACT WITH ENVIRONMENT
LIGHT OILS STORAGE
MIDDLE OILS STORAGE
HEAVY OILS STORAGE
NAPHTHA PRODUCT STORAGE
LURGI OIL RECOVERY FIGURE 3.3-4
NAPHTHA RECOVERY FIGURE 3.3-5
GAS COMPRESSION FIGURE 3.3-6
DEA/TEG TREATMENT FIGURE 3.3-7
STRETFORD SULFUR RECOVERY FIGURE 3.3-8
DRIED FUEL GAS TO PIPELINE
LIQUID SULFUR STORAGE
LURGI-RUHRGAS RETORTING AND FLUE GAS DISCHARGE (ELECTROSTATIC PRECIPITATOR) FIGURE 3.3-3
AMMONIA RECOVERY FIGURE 3.3-9
ANHYDROUS AMMONIA STORAGE
SOLID STATE DISPOSAL (PIT BACKFILL) FIGURE 3.3-10
MINING, CRUSHING & STORAGE FIGURE 3.3-2
WATER MANAGEMENT AND AERATION POND FIGURE 3.3-11
CW
BOXES WITH HEAVY LINES INDICATE POLLUTION CONTROL ACTIVITIES

SOURCE DRI based on Gulf Oil Corp. and Standard Oil Co. (Indiana), March 1976, and Rio Blanco Oil Shale Co., February 1981

FIGURE 3.3-1 OVERALL PLANT COMPLEX

The individual, unit process flow diagrams are presented throughout this section (see Figures 3.3-2 through 3.3-11); also, Figure 3.3-12 provides details on the water management system for the entire plant complex. In each diagram, streams enter on the left and exit on the right, and mass flows are given at the bottom. Composition data on major process and waste streams can be found in Section 4.

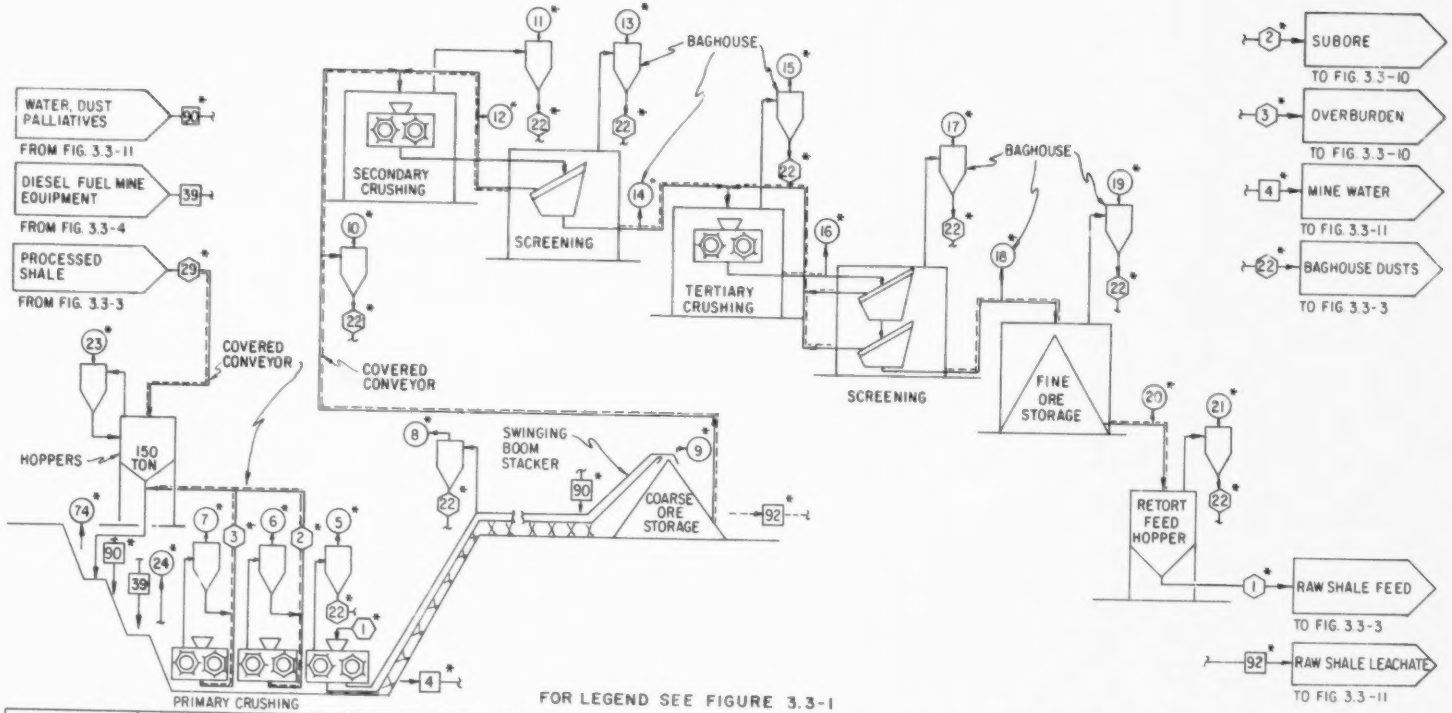
3.3.1 Mining, Crushing and Transport of Raw Shale

A flow diagram illustrating mining and crushing processes for the Lurgi-Open Pit oil shale complex is presented in Figure 3.3-2.

Run-of-mine oil shale, overburden, and subore will be reduced to 6 to 8 inches in size in separate 72-inch gyratory crushers located in the pit. These primary crushers will have the capability of being moved as the pit migrates. Airborne particulate matter from the primary crushing operations will be controlled with baghouses and the uncontrolled particulates will be emitted to the atmosphere (streams 5, 6 and 7). The baghouse dust collected from the overburden and subore crushing operations will be combined with the crushed materials, while the shale dust (stream 22) will be added into the feed to the Lurgi-Ruhrgas retorts. The primary crushed shale will be conveyed to the coarse ore stockpile located on the surface and subsequently reduced in size by secondary and tertiary crushing. All crushing, screening, and transfer operations will use baghouses to control dust emissions. Fugitive dust associated with the mining and crushing operations will be controlled with water sprays and other dust palliatives (stream 90). The fine crushed shale will be stockpiled in an enclosed storage bin and conveyed to the retort feed hopper. The airborne particulates from the storage bin, conveyor, and feed hopper will be controlled with baghouses and then emitted to the atmosphere (streams 19, 20 and 21, respectively). Individual conveyors will then transport the shale from the feed hopper to each of the retorts. Fugitive dust from these conveyors will also be controlled with baghouses. (Alternately, the shale can be distributed to the retorts directly from the storage bin. In such a case, the feed hopper depicted in the figure would not be included; therefore, stream 21 would not exist. Instead, each retort would use an individual feed hopper, which would be controlled with the baghouse installed on the feed conveyor.) The crushed overburden and subore (streams 2 and 3) and processed shale (stream 29) will be transferred by covered conveyor to 150-ton truck load-out hoppers for redistribution in the back of the pit.

Due to the interruption of the existing aquifers during development of the pit, a considerable amount of water will need to be pumped through the dewatering wells. Shown as stream 4 on the flow diagram, this water will be sent to the water management system (see Section 3.3.10) for clarification and subsequent treatment before being used throughout the plant complex.

Leachate from the raw shale pile, if present, and storm runoff will be pumped to an oil/water separator and then used for processed shale moisturizing. The various mining and disposal equipment (e.g., power shovels, trucks, crushers) operates on diesel fuel. The exhaust gases


WATER, DUST PALLIATIVES
FROM FIG. 3.3-11
90
DIESEL FUEL MINE EQUIPMENT
FROM FIG. 3.3-4
39
PROCESSED SHALE
FROM FIG. 3.3-3
29
23
COVERED CONVEYOR
HOPPERS
150 TON
74
90
39
24
7
3
6
2
5
22
1
4
PRIMARY CRUSHING
10
22
COVERED CONVEYOR
SECONDARY CRUSHING
12
11
22
13
22
BAGHOUSE
SCREENING
14
15
22
TERTIARY CRUSHING
8
22
SWINGING BOOM STACKER
90
9
COARSE ORE STORAGE
92
16
SCREENING
17
22
BAGHOUSE
18
19
22
FINE ORE STORAGE
20
21
22
RETORT FEED HOPPER
1
RAW SHALE FEED
TO FIG. 3.3-3
92
RAW SHALE LEACHATE
TO FIG 3.3-11
2
SUBORE
TO FIG. 3.3-10
3
OVERBURDEN
TO FIG. 3.3-10
4
MINE WATER
TO FIG. 3.3-11
22
BAGHOUSE DUSTS
TO FIG 3.3-3

FOR LEGEND SEE FIGURE 3.3-1

STREAM NO.	1*	2*	3*	4*	5*	6*	7*	8*	9*	10*	11*	12*	13*	14*	15*	16*	17*	18*	19*	20*	21*	22*	23*	24*	29*	39	74*	90*	92*
STREAM IDENTITY	RAW SHALE FEED	SUB-ORE	OVER-BURDEN	MINE WATER	← BAGHOUSE EMISSIONS →																	BAG-HOUSE DUSTS	BAG-HOUSE EMISS.	DIESEL EMISS.	PROC. SHALE	DIESEL FUEL MINE EQUIP.	FUGI-TIVE DUSTS	WATER DUST PALL.	RAW SHALE LEACH.
FLOW RATE 10³ ACFM 10³ lb/hr gpm	9799	992	5175	16,500	122.2 15.7 lb/hr	12.2 1.6 ←	63.8 8.2	72.6 0.4	121.5 15.6	40.4 0.2	558.4 71.9	40.4 0.2	558.4 71.9	40.4 0.2	628.4 80.8	80.8 0.4	628.2 80.8	40.4 0.2	28.8 3.7	40.4 0.2 →	212.4 27.3 lb/hr	118.1	64.5 8.3 lb/hr	151.3	9733	N.D.	52.0 lb/hr	1568	N.D.
TEMP, °F	AMB	←																				→	AMB	N.D.	180	AMB	AMB	AMB	AMB
PRESSURE, psig	AMB	←																										→	AMB

SOURCE: DRI based on Gulf Oil Corp. and Standard Oil Co (Indiana), March 1976, and Rio Blanco Oil Shale Co., February 1981

FIGURE 3.3-2 MINING & CRUSHING

(stream 24) from this equipment will be controlled through the use of catalytic converters.

3.3.2 Lurgi-Ruhrgas Aboveground Retorting

The Lurgi-Ruhrgas aboveground retorting process is shown in Figure 3.3-3. Raw shale (stream 1) from the crushers and collected dust (stream 22) from the baghouses provide the feed to the screw mixer where pyrolysis occurs. Vapors containing retort gases, oil mist, water vapor, and some processed shale particulates exit the screw mixer and pass through two cyclones. Processed shale particulates are removed in the cyclones and the vapors (stream 26) continue on to the oil recovery system.

Processed shale exits the screw mixer into a surge vessel where it combines with particulates captured by the gas stream cyclones. The processed shale is then forced up a lift pipe by injection of preheated air (stream 25). A portion of the retort gas (stream 35) is a necessary addition to the lift pipe to sustain combustion of the residual organic matter on the processed shale. Oily dust from the oil/dust centrifuge (stream 43) is also injected into the bottom of the lift pipe. Combustion of residual organic matter on processed shale particles and oil from oily dust produce flue gas and heat.

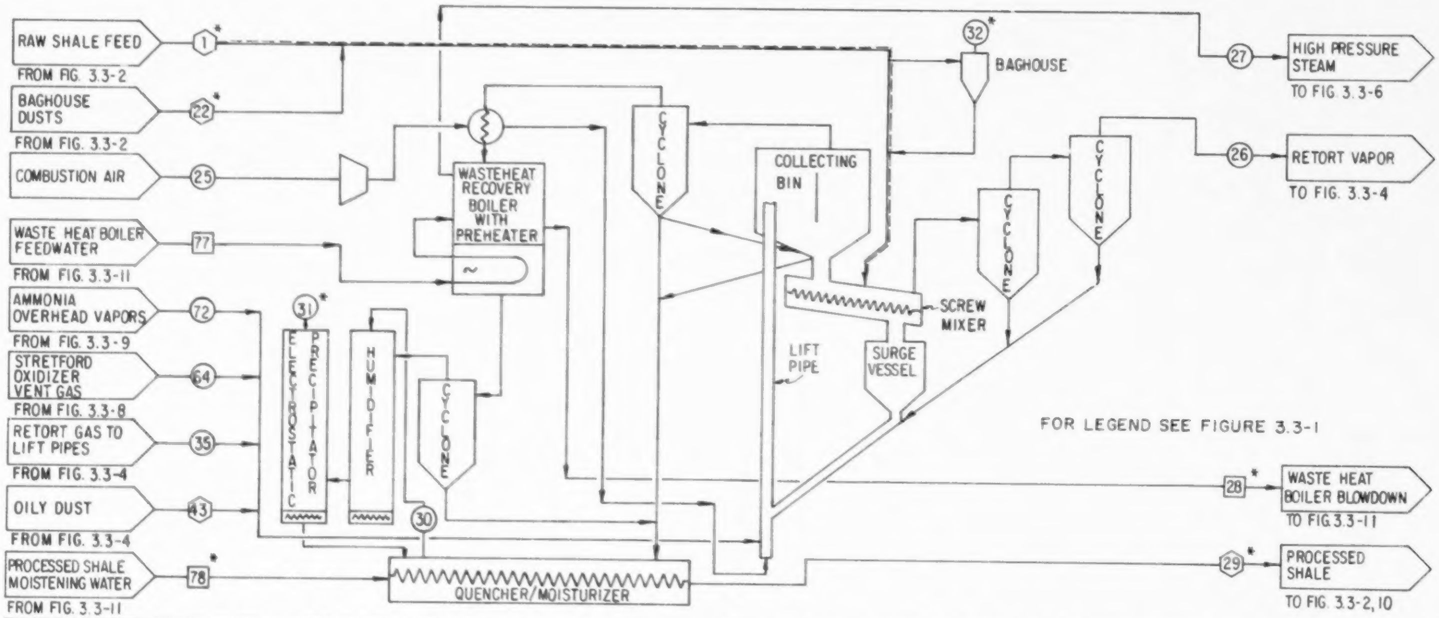
The processed shale particles then enter a collecting bin which recycles a predetermined amount of hot processed shale into the screw mixer to provide heat necessary to raise the raw shale feed to pyrolysis temperature. The processed shale is mixed with raw shale in the screw mixer in a mass ratio of approximately 6:1 (Marnell, September 1976; Schmalfeld, July 1975).

Hot flue gas and entrained processed shale particles exit the collecting bin and enter a cyclone where most particulates are removed and fed to a processed shale quencher/moisturizer. The flue gas then enters a waste heat recovery boiler where high pressure steam (stream 27) is produced through heat transfer to the entering waste heat boiler feedwater (stream 77). The high pressure steam is utilized as the prime mover for the turbine-driven compressors in the gas compression section (Figure 3.3-6).

The flue gas continues through another cyclone for further particulate removal and then enters a humidifier and an electrostatic precipitator and is vented to the atmosphere as stream 31. Processed shale quenching and moistening water (stream 78) enters the quencher/moisturizer where the processed shale is cooled to below 200°F and is wetted to contain approximately 19% water by weight. The moisturized processed shale (stream 29) is then sent for disposal. The steam produced by the quenching operation is added to the flue gas via the humidifier. This aids in the electrostatic precipitation of the particulates.

3.3.3 Lurgi-Ruhrgas Oil Recovery

The Lurgi-Ruhrgas oil recovery system, shown in Figure 3.3-4, has three stages involving two hot oil scrubbers and one cool water scrubber. The oil


RAW SHALE FEED
FROM FIG. 3.3-2
BAGHOUSE DUSTS
FROM FIG. 3.3-2
COMBUSTION AIR
WASTE HEAT BOILER FEEDWATER
FROM FIG. 3.3-11
AMMONIA OVERHEAD VAPORS
FROM FIG. 3.3-9
STRETFORD OXIDIZER VENT GAS
FROM FIG. 3.3-8
RETORT GAS TO LIFT PIPES
FROM FIG. 3.3-4
OILY DUST
FROM FIG. 3.3-4
PROCESSED SHALE MOISTENING WATER
FROM FIG. 3.3-11
1*
22*
25
77
72
64
35
43
78*
31*
30
32*
27
26
28*
29*
WASTEHEAT RECOVERY BOILER WITH PREHEATER
ELECTROSTATIC PRECIPITATOR
HUMIDIFIER
CYCLONE
CYCLONE
CYCLONE
CYCLONE
COLLECTING BIN
LIFT PIPE
SCREW MIXER
SURGE VESSEL
BAGHOUSE
QUENCHER/MOISTURIZER
HIGH PRESSURE STEAM
TO FIG. 3.3-6
RETORT VAPOR
TO FIG. 3.3-4
WASTE HEAT BOILER BLOWDOWN
TO FIG. 3.3-11
PROCESSED SHALE
TO FIG. 3.3-2, 10
FOR LEGEND SEE FIGURE 3.3-1

STREAM NO.	① *	㉒ *	㉕	㉖	㉗	28 *	㉙ *	㉚	㉛ *	㉜ *	㉟	43	64	72	77	78 *
STREAM IDENTITY	RAW SHALE FEED	BAGHOUSE DUSTS	COMBUSTION AIR	RETORT VAPORS	HIGH PRESSURE STEAM	WASTE HEAT BOILER BLOWDOWN	PROCESSED SHALE	STEAM TO HUMIDIFIER	FLUE GAS	RETORT FEED BAGHOUSE EMISSION	RETORT GAS TO LIFT PIPES	OILY DUST	STRETFORD OXIDIZER VENT GAS	AMMONIA OVERHEAD VAPORS	WASTE HEAT BOILER FEEDWATER	PROCESSED SHALE MOISTENING WATER
FLOW RATE: 10^3 ACFM 10^3 lb/hr gpm	9799	118.1	5439	1425.2	1060	21	9733	992	3820 7202.6	239.2 1.2 lb/hr	107.0	78.6	24.3	20.2	2120	5624
TEMP, °F	AMB	AMB	740	950	475	N.D.	180	N.D	460	AMB	90	N.D.	N.D.	210	AMB	AMB
PRESSURE, psig	AMB	AMB	N.D.	AMB	550	AMB	AMB	N.D.	AMB	AMB	AMB	AMB	AMB	3	AMB	AMB

SOURCE: DRI based on Schmalfeld, July 1975

FIGURE 3.3-3 LURGI-RUHRGAS RETORTING

recovery system primarily removes oil mists and water vapor from the entering retort vapors (stream 26).

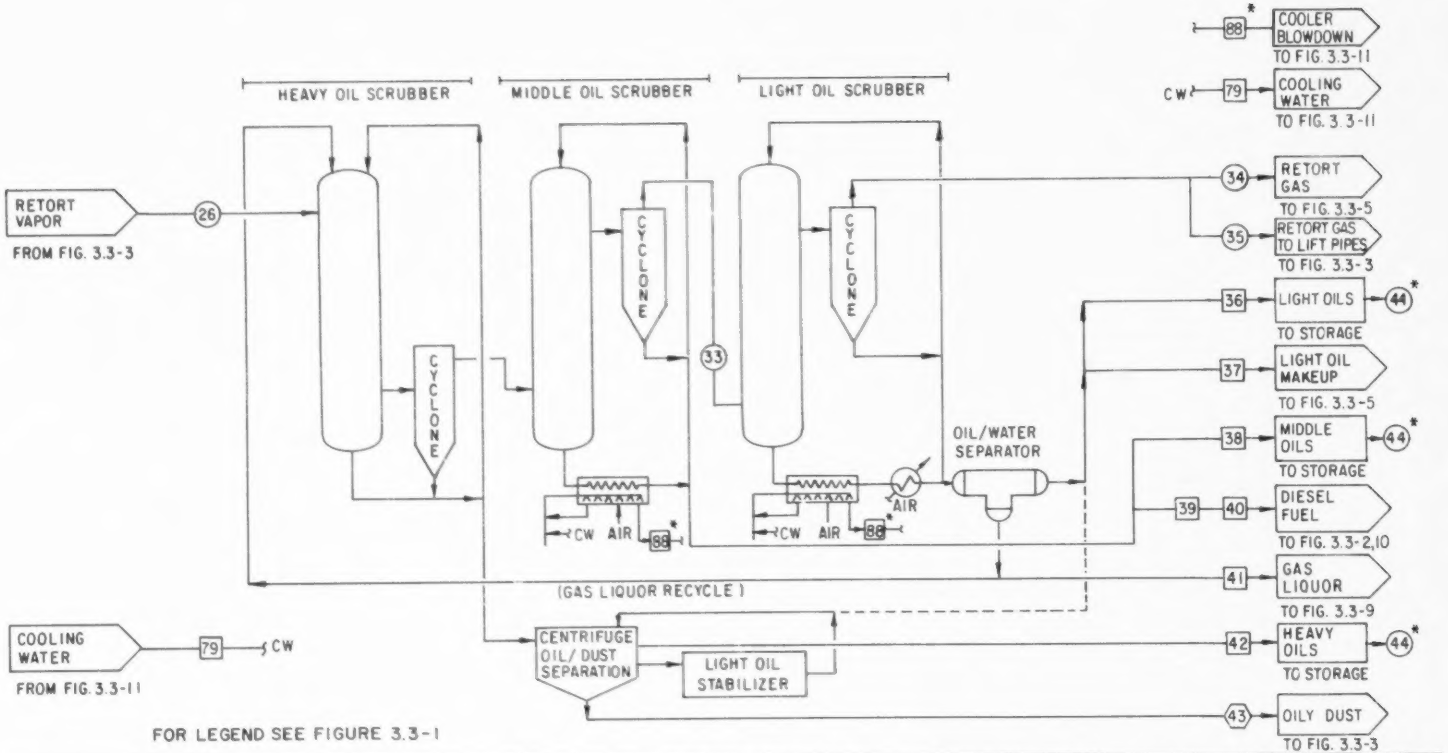
The first oil scrubber removes heavy oils and particulate material from the gas stream but retains water in the vapor phase due to the high temperature involved. Gas liquor, which is obtained in the latter part of the oil recovery system, is recycled to the heavy oil scrubber to decrease the temperature of the vapors through water evaporation. Dusty heavy oils exit at the bottom of the scrubber at approximately 350°F and enter a centrifuge for dust/oil separation. A small stream of light oils is a necessary addition to the centrifugation process in order to thin the highly viscous heavy oils, thereby enabling a more effective separation. Centrifugation can be a two-stage process, coupled with solids drying and light oil stabilization processes (Rio Blanco Oil Shale Co., February 1981). Oily dust (stream 43) recovered through centrifugation is recycled to the Lurgi retort lift pipe. The dust-free heavy oils (stream 42) recovered through centrifugation are pumped to storage. Retort gas exits the first scrubber and passes through a cyclone for removal of oil droplets and particulates before entering the second oil scrubber for middle oils removal.

The second oil scrubber operates similar to the first, affording removal of middle oils from the retort gas. This scrubber operates at a temperature lower than the first, yet above the dew point of the gas so that moisture condensation does not occur; therefore, the middle oils (stream 38) recovered from this unit are free of water and may be pumped directly to storage. The operating temperature for the scrubber is about 150°F, and it is controlled with humidified air coolers.

The third scrubber operates at a temperature low enough to promote condensation of light oil vapors and moisture. This unit recirculates a portion of the condensed oil/water mixture to aid in scrubbing of oil and to promote removal of ammonia. The exit temperature of this scrubber is approximately 90°F. Light oils (stream 36) and gas liquor (stream 41) condensed from this scrubber undergo separation, with light oils being pumped to storage and gas liquor continuing on to the ammonia recovery unit (see Section 3.3.8). The oil storage tanks will be a source of fugitive hydrocarbon emissions (stream 44).

3.3.4 Lurgi Lean Oil Absorber and Naphtha Stripper

The lean oil absorber and naphtha stripper unit, shown in Figure 3.3-5, fractionates the retort gas (stream 34) after oil recovery into naphtha (stream 46) and noncondensable hydrocarbons such as C_1's, C_2's and C_3's. This unit is used to absorb the naphtha from the gas into a naphtha-free or lean oil. Noncondensable hydrocarbons are not absorbed in the lean oil and exit the absorber overhead as naphtha-free retort gas (stream 45) for further cleanup and use. Naphtha is then stripped from the naphtha-rich oil and condensed for collection and storage, while the stripped lean oil is recycled to the absorber. Makeup lean oil (stream 37) is obtained, as required, from light oils produced in the oil recovery unit.

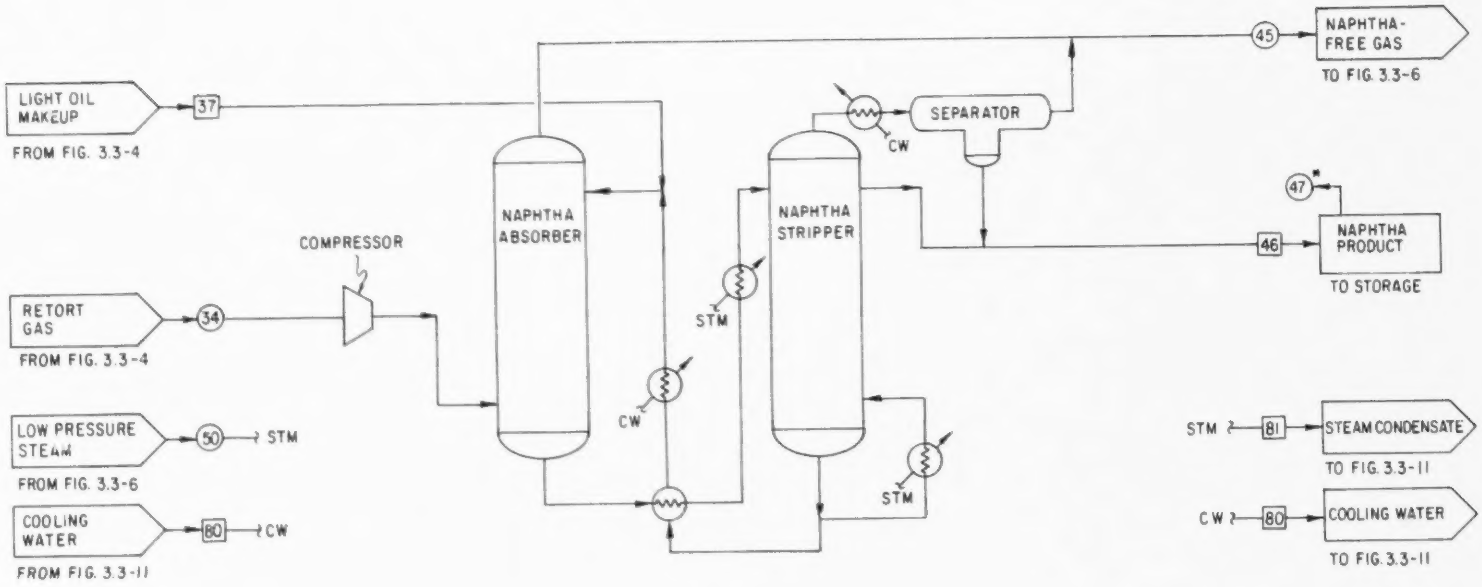

HEAVY OIL SCRUBBER
MIDDLE OIL SCRUBBER
LIGHT OIL SCRUBBER
RETORT VAPOR
FROM FIG. 3.3-3
26
CYCLONE
CYCLONE
CYCLONE
33
CW
AIR
88
OIL/WATER SEPARATOR
(GAS LIQUOR RECYCLE)
COOLING WATER
FROM FIG. 3.3-11
79
CENTRIFUGE OIL/DUST SEPARATION
LIGHT OIL STABILIZER
COOLER BLOWDOWN
TO FIG. 3.3-11
COOLING WATER
TO FIG. 3.3-11
34
RETORT GAS
TO FIG. 3.3-5
35
RETORT GAS TO LIFT PIPES
TO FIG. 3.3-3
36
LIGHT OILS
TO STORAGE
44
37
LIGHT OIL MAKEUP
TO FIG. 3.3-5
38
MIDDLE OILS
TO STORAGE
39
40
DIESEL FUEL
TO FIG. 3.3-2,10
41
GAS LIQUOR
TO FIG. 3.3-9
42
HEAVY OILS
TO STORAGE
43
OILY DUST
TO FIG. 3.3-3

FOR LEGEND SEE FIGURE 3.3-1

STREAM NO.	㉖	㉝	㉞	㉟	36	37	38	39	40	41	42	43	㊹*	79	88*
STREAM IDENTITY	RETORT VAPOR	RAW RETORT GAS	RETORT GAS	RETORT GAS TO LIFT PIPES	LIGHT OILS TO STORAGE	LIGHT OIL MAKEUP TO NAPHTHA RECOVERY	MIDDLE OILS TO STORAGE	DIESEL FUEL MINE EQUIPMENT	DIESEL FUEL DISPOSAL EQUIPMENT	GAS LIQUOR	HEAVY OILS TO STORAGE	OILY DUST	HC EMISS. FROM STORAGE TANKS	COOLING WATER	COOLER BLOWDOWN
FLOW RATE: 10^3 ACFM 10^3 lb/hr gpm	1425.2	738.8	149.0	107.0	185.2 406	N.D.	411.3 937	N.D.	N.D.	297.7 586	196.5 416	78.6	(includes #47) 65.5 lb/hr	325	661
TEMP., °F	950	150	90	90	90	90	150	AMB	AMB	90	350	N.D.	AMB	AMB	N.D.
PRESSURE, psig	AMB ←														→ AMB

SOURCE: DRI based on Rio Blanco Oil Shale Co., February 1981

FIGURE 3.3-4 OIL RECOVERY


LIGHT OIL MAKEUP
FROM FIG. 3.3-4
37
RETORT GAS
FROM FIG. 3.3-4
34
COMPRESSOR
LOW PRESSURE STEAM
FROM FIG. 3.3-6
50
STM
COOLING WATER
FROM FIG. 3.3-11
80
CW
NAPHTHA ABSORBER
CW
NAPHTHA STRIPPER
STM
STM
CW
SEPARATOR
45
NAPHTHA-FREE GAS
TO FIG. 3.3-6
47*
46
NAPHTHA PRODUCT
TO STORAGE
STM
81
STEAM CONDENSATE
TO FIG. 3.3-11
CW
80
COOLING WATER
TO FIG. 3.3-11

FOR LEGEND SEE FIGURE 3.3-1

STREAM NO.	(34)	37	(45)	46	(47)*	(50)	80	81
STREAM IDENTITY	RETORT GAS	LIGHT OIL MAKEUP	NAPHTHA-FREE GAS	NAPHTHA PRODUCT TO STORAGE	HC EMISSIONS FROM NAPHTHA STORAGE	L. P. STEAM	COOLING WATER	STEAM CONDENSATE
FLOW RATE: 10^3 ACFM 10^3 lb/hr gpm	149.0	N.D.	121.7	27.3	(included in #44) N.D.	10	5	20
TEMP, °F	90	90	AMB	AMB	AMB	275	AMB	N.D.
PRESSURE, psig	AMB	AMB	AMB	N.D.	AMB	60	AMB	AMB

SOURCE: DRI based on Rio Blanco Oil Shale Co., February 1981

FIGURE 3.3-5 NAPHTHA RECOVERY

3.3.5 Retort Gas Compression and Cooling

Figure 3.3-6 shows a flow diagram for retort gas compression and cooling. This operation consists of three-stage compression of the retort gas from ambient pressure to about 1,000 psig. This step also serves to eliminate a considerable amount of moisture and ammonia from the retort gas.

The naphtha-free retort gas (stream 45) from the naphtha stripper enters the first of four flash drums. This flash drum operates at ambient pressure and also receives the compressed condensate streams from the other three drums. The dissolved gases in the compressed condensates are flashed in the first drum and combined with the retort gas. The condensate (stream 49) is recovered at the bottom and sent to the ammonia recovery unit (Section 3.3.8) along with the gas liquor. The combined gas stream from the first flash drum then enters the first stage of compression. The compressor discharge is water cooled and fed to the second flash drum, then to the second compression stage, and so on. Eventually, the compressed retort gas (stream 48) is obtained from the last of the flash drums and transported to the diethanolamine (DEA) absorber for the removal of acid gases (Figure 3.3-7).

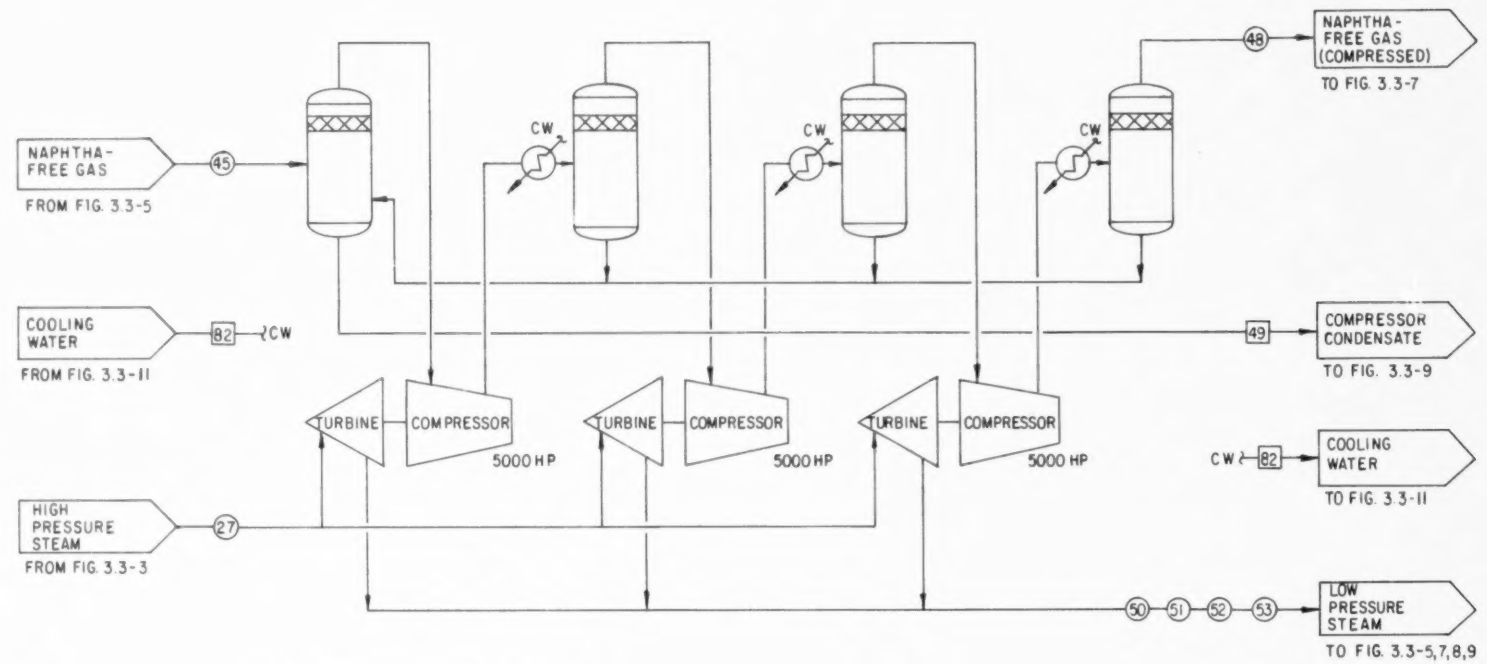
The compressors are driven by steam turbines using high pressure steam (stream 27) from the waste heat boiler. The turbines are of a noncondensing type, discharging low pressure steam (streams 50, 51, 52 and 53) to be used in other plant operations.

3.3.6 Amine Treatment/Triethylene Glycol Dehydration

The amine treatment and dehydration for the compressed retort gas are shown in Figure 3.3-7. The compressed naphtha-free gas (stream 48) enters the gas treating column and is scrubbed with 30% by weight DEA (diethanolamine/water solution) to remove H_2S. The CO_2 level in the gas stream is also reduced to a low level by the amine solution. The rich amine solution leaving the absorber is regenerated by steam stripping, which produces the acid gases (stream 58) from the top of the amine regenerator. This stream is sent to the Stretford unit where the hydrogen sulfide is converted and recovered as elemental sulfur. The retort gas (stream 56) emerging from the top of the DEA absorber is virtually free of acid gases and enters the triethylene glycol (TEG) dehydrating system. The retort gas is scrubbed with the glycol, which picks up the residual moisture in the gas. The dry gas (stream 57) is sent to the pipeline, and the glycol solution is regenerated by steam stripping and then recycled. The TEG regenerator overhead vapor (stream 60), containing mostly steam with a slight amount of glycol, is emitted to the atmosphere.

3.3.7 Stretford Sulfur Process

A flow diagram for the Stretford process is shown in Figure 3.3-8. This process affords simultaneous removal and recovery of hydrogen sulfide from the gaseous feeds containing low amounts of H_2S. High concentrations of H_2S as well as CO_2 are detrimental to the efficiency of the process.

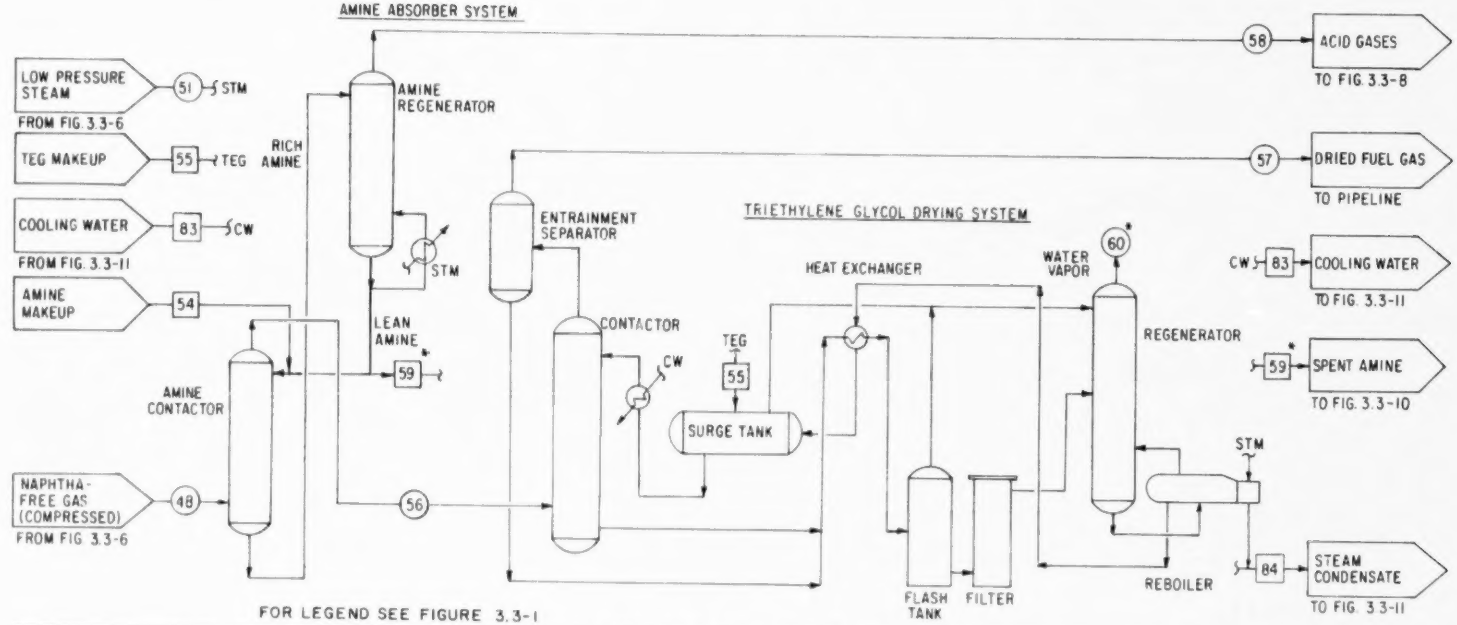

NAPHTHA-FREE GAS
FROM FIG. 3.3-5
45
COOLING WATER
FROM FIG. 3.3-11
82
CW
HIGH PRESSURE STEAM
FROM FIG. 3.3-3
27
CW
CW
CW
TURBINE
COMPRESSOR
5000 HP
TURBINE
COMPRESSOR
5000 HP
TURBINE
COMPRESSOR
5000 HP
48
NAPHTHA-FREE GAS (COMPRESSED)
TO FIG. 3.3-7
49
COMPRESSOR CONDENSATE
TO FIG. 3.3-9
CW
82
COOLING WATER
TO FIG. 3.3-11
50
51
52
53
LOW PRESSURE STEAM
TO FIG. 3.3-5,7,8,9

FOR LEGEND SEE FIGURE 3.3-1

STREAM NO.	㉗	㊺	㊽	49	㊿	(51)	(52)	(53)	82
STREAM IDENTITY	HIGH PRESSURE STEAM	NAPHTHA-FREE GAS	NAPHTHA-FREE GAS (COMPRESSED)	COMPRESSOR CONDENSATE	L.P. STEAM TO NAPHTHA RECOVERY	L.P. STEAM TO DEA UNIT	L.P. STEAM TO STRETFORD UNIT	L.P. STEAM TO AMMONIA RECOVERY	COOLING WATER
FLOW RATE: 10^3 ACFM 10^3 lb/hr gpm	1060	121.7	118.0	3.8 8	10	130	1	53	8
TEMP., °F	475	AMB	AMB	N.D.	275	275	275	275	AMB
PRESSURE, psig	550	AMB	1000	N.D.	60	60	60	60	AMB

SOURCE: DRI based on information provided by SWEC

FIGURE 3.3-6 RETORT GAS COMPRESSION AND COOLING

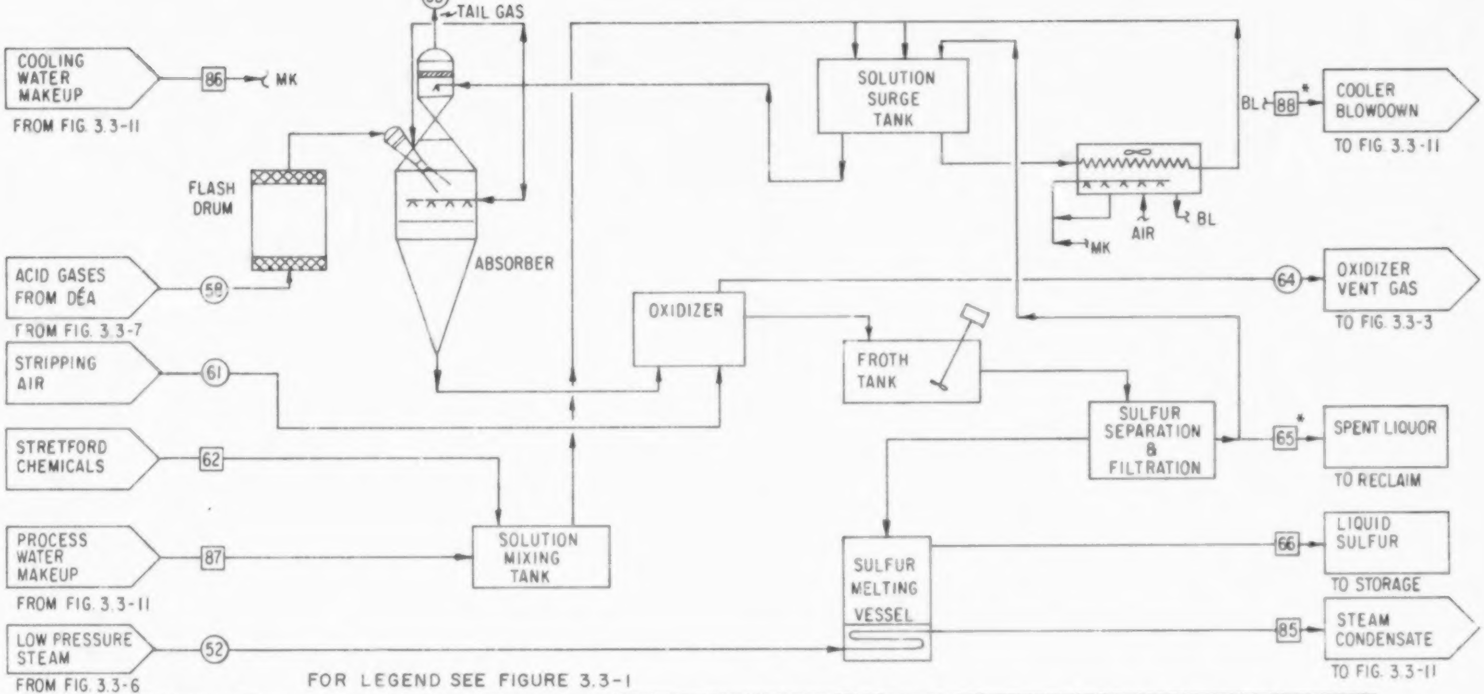

AMINE ABSORBER SYSTEM
LOW PRESSURE STEAM
FROM FIG. 3.3-6
(51) STM
TEG MAKEUP
55 TEG
COOLING WATER
FROM FIG. 3.3-11
83 CW
AMINE MAKEUP
54
NAPHTHA-FREE GAS (COMPRESSED)
FROM FIG. 3.3-6
(48)
AMINE CONTACTOR
RICH AMINE
AMINE REGENERATOR
STM
LEAN AMINE
59 *
(56)
ENTRAINMENT SEPARATOR
CONTACTOR
CW
TEG
55
SURGE TANK
TRIETHYLENE GLYCOL DRYING SYSTEM
HEAT EXCHANGER
FLASH TANK
FILTER
WATER VAPOR
(60) *
REGENERATOR
REBOILER
STM
(58)
ACID GASES
TO FIG. 3.3-8
(57)
DRIED FUEL GAS
TO PIPELINE
CW 83
COOLING WATER
TO FIG. 3.3-11
59 *
SPENT AMINE
TO FIG. 3.3-10
84
STEAM CONDENSATE
TO FIG. 3.3-11

FOR LEGEND SEE FIGURE 3.3-1

STREAM NO.	(48)	(51)	54	55	(56)	(57)	(58)	59 *	(60) *	83	84
STREAM IDENTITY	NAPHTHA-FREE GAS (COMPRESSED)	L.P. STEAM	AMINE MAKEUP	TEG MAKEUP	SWEET GAS	DRIED FUEL GAS TO PIPELINE	ACID GASES	SPENT AMINE	TEG REGENERATION VENT EMISSION	COOLING WATER	STEAM CONDENSATE
FLOW RATE: 10^3 ACFM 10^3 lb/hr gpm	118.0	130	N.D.	N.D.	61.9	61.8	10.5 57.1	N.D.	N.D.	66	260
TEMP, °F	AMB	275	AMB	AMB	N.D.	AMB	105	N.D.	AMB	AMB	N.D.
PRESSURE, psig	1000	60	AMB	AMB	1000	N.D.	AMB	AMB	AMB	AMB	AMB

SOURCE: DRI based on information provided by SWEC

FIGURE 3.3-7 DIETHANOLAMINE/TRIETHYLENE GLYCOL TREATMENT


COOLING WATER MAKEUP
FROM FIG. 3.3-11
86
MK
FLASH DRUM
ACID GASES FROM DEA
FROM FIG. 3.3-7
58
STRIPPING AIR
61
STRETFORD CHEMICALS
62
PROCESS WATER MAKEUP
FROM FIG. 3.3-11
87
LOW PRESSURE STEAM
FROM FIG. 3.3-6
52
63*
TAIL GAS
ABSORBER
SOLUTION MIXING TANK
OXIDIZER
SOLUTION SURGE TANK
FROTH TANK
SULFUR SEPARATION & FILTRATION
SULFUR MELTING VESSEL
MK
AIR
BL
BL
88*
COOLER BLOWDOWN
TO FIG. 3.3-11
64
OXIDIZER VENT GAS
TO FIG. 3.3-3
65*
SPENT LIQUOR
TO RECLAIM
66
LIQUID SULFUR
TO STORAGE
85
STEAM CONDENSATE
TO FIG. 3.3-11

FOR LEGEND SEE FIGURE 3.3-1

STREAM NO.	52	58	61	62	63*	64	65*	66	85	86	87	88*
STREAM IDENTITY	L.P. STEAM	ACID GASES FROM DEA	STRIPPING AIR	STRETFORD CHEMICALS	STRETFORD TREATED ACID GASES	OXIDIZER VENT GAS	SPENT LIQUOR TO RECLAIM	LIQUID SULFUR PRODUCT	STEAM CONDENSATE	COOLING WATER MAKEUP	PROCESS WATER MAKEUP	COOLER BLOWDOWN
FLOW RATE: 10^3 ACFM		10.5			9.0	6.0						
10^3 lb/hr	1	57.1	20.9	0.03	53.0	24.3	N.D.	0.7				
gpm								(7.6 LTPSD)	2	2	3	661
TEMP, °F	275	105	AMB	AMB	95	N.D.	95	260	N.D.	AMB	AMB	N.D.
PRESSURE, psig	60	AMB ←										→ AMB

SOURCE: DRI based on Peabody Process Systems, Inc., February 1981

FIGURE 3.3-8 STRETFORD SULFUR RECOVERY

The Stretford process consists of H_2S absorption, solution regeneration and sulfur recovery systems. The Stretford solution consists of a buffered solution of sodium carbonates, anthraquinone disulfonic acid (ADA), and sodium vanadate which, in effect, oxidize H_2S to elemental sulfur. The reactants are regenerated by stripping and oxidizing with air, then are recycled.

The acid gases from the amine regenerator (stream 58) are introduced into the absorber through venturi inlets under the solution level. By reacting with vanadate in the presence of ADA, H_2S is converted to elemental sulfur, which then floats to the surface and is skimmed off in the oxidizer. After filtering and melting, the sulfur product (stream 66) is taken to storage. The treated acid gases (stream 63) are released to the atmosphere without further treatment.

Stripping air is purged through the Stretford solution in the oxidizer tank to regenerate the ADA. The oxidizer vent gas, containing the stripping air with some desorbed materials (stream 64), is then used as a combustion air source for the lift pipes. The regenerated solution is recycled to the absorbers. Some nonregenerable compounds like thiosulfates form during the solution regeneration. These are removed periodically as part of the spent liquor (stream 65), which is sent for reclaim.

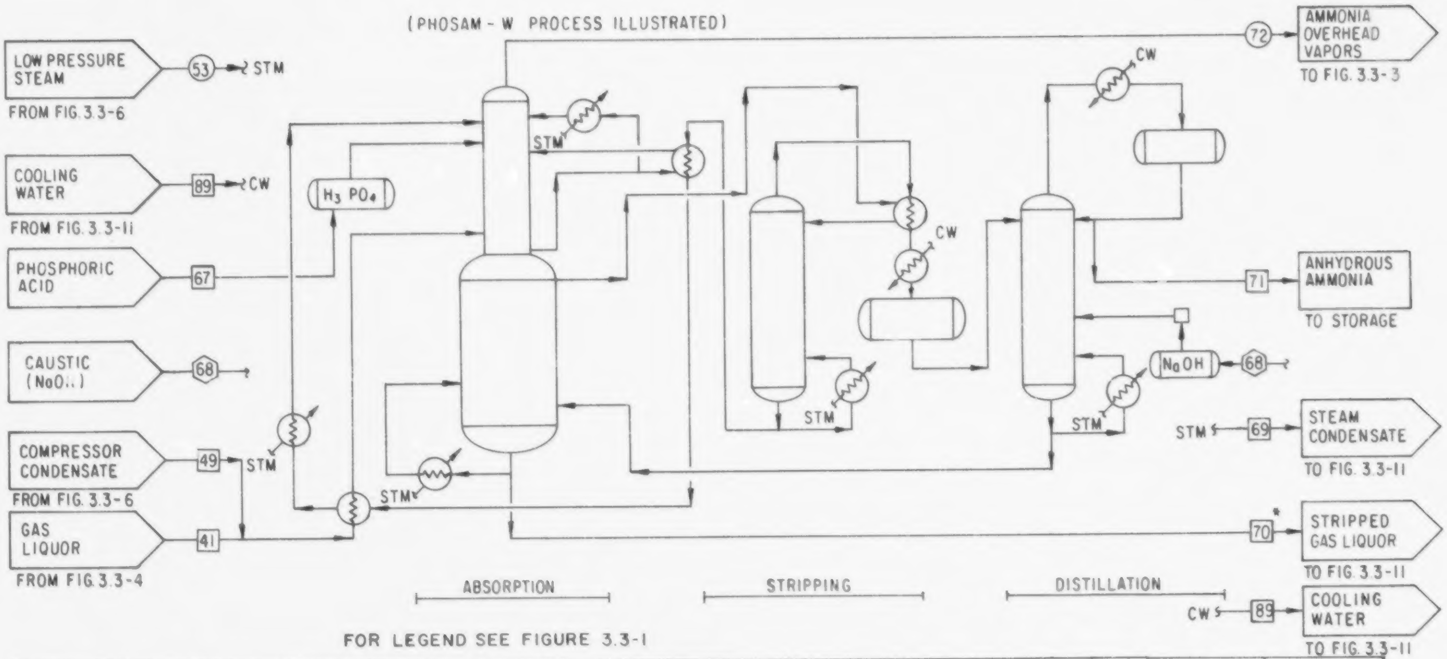
3.3.8 Ammonia Recovery Process

A schematic flow diagram for an ammonia recovery process is presented in Figure 3.3-9. This unit treats combined ammoniacal gas liquors (streams 41 and 49) from the oil recovery and gas compression units, respectively, for recovery of anhydrous ammonia.

The ammonia recovery process illustrated consists of a water stripper, an ammonia absorber, an ammonia stripper, and an ammonia concentrator or boiler. The gas liquor feed is introduced to the water stripper in which the dissolved ammonia and other volatile matter are evolved by steam stripping the water. Sodium hydroxide may be added to the aqueous charge to facilitate release of fixed ammonia. The stripped water (stream 70) is used in processed shale moisturizing.

In the ammonia absorber, the ammonia released from the gas liquor is absorbed out of the vapor phase in a phosphoric acid solution. A solution stoichiometry between monoammonium phosphate and diammonium phosphate is maintained for efficient absorption of ammonia. Unabsorbed gases such as H_2S and CO_2 continue on, as the ammonia recovery unit overhead vapors (stream 72), to the Lurgi retort lift pipes for incineration of H_2S.

Desorption of the ammonia from the ammonium phosphate solution takes place in the ammonia stripper section. Both temperature and pressure are increased and steam is passed through the solution. An aqueous solution of 10-20% ammonia is condensed overhead, while the stripped or lean solution is recycled to the absorption section. Ammonia is then obtained in an anhydrous state (stream 71) in the distillation section by steam stripping the aqueous ammonia solution and fractionating the vapors.


(PHOSAM - W PROCESS ILLUSTRATED)
LOW PRESSURE STEAM
FROM FIG. 3.3-6
53
STM
COOLING WATER
FROM FIG. 3.3-11
89
CW
PHOSPHORIC ACID
67
H3 PO4
CAUSTIC (NaOH)
68
COMPRESSOR CONDENSATE
FROM FIG. 3.3-6
49
STM
GAS LIQUOR
FROM FIG. 3.3-4
41
STM
STM
CW
STM
CW
STM
NaOH
68
72
AMMONIA OVERHEAD VAPORS
TO FIG. 3.3-3
71
ANHYDROUS AMMONIA
TO STORAGE
STM
69
STEAM CONDENSATE
TO FIG. 3.3-11
70*
STRIPPED GAS LIQUOR
TO FIG. 3.3-11
CW
89
COOLING WATER
TO FIG. 3.3-11
ABSORPTION
STRIPPING
DISTILLATION

FOR LEGEND SEE FIGURE 3.3-1

STREAM NO		41	49	53	67	68	69	70*	71	72	89
STREAM IDENTITY		GAS LIQUOR	COMPRESSOR CONDENSATE	LOW PRESSURE STEAM	PHOSPHORIC ACID	CAUSTIC (NaOH)	STEAM CONDENSATE	STRIPPED GAS LIQUOR	ANHYDROUS AMMONIA STORAGE	AMMONIA OVERHEAD VAPORS	COOLING WATER
FLOW RATE:	10^3 ACFM									7.6	
	10^3 lb/hr	297.7	3.8	53	0.01	0.3		279.7	1.9	20.2	
	gpm	586	8				106	558	22.6 TPSD		1080
TEMP, °F		90	N.D.	275	AMB	AMB	N.D.	180	N.D.	210	AMB
PRESSURE, psig		AMB	N.D.	60	AMB	AMB	AMB	6	N.D.	3	AMB

SOURCE: DRI based on U.S.S. Engineers and Consultants, Inc., April 1978

FIGURE 3.3-9 AMMONIA RECOVERY

3.3.9 Solid Waste Disposal

Figure 3.3-10 presents a conceptual design for solid waste disposal via backfilling the open pit. This disposal approach, which was mentioned in the original DDP for Tract C-a, was to commence after 30 years of pit development, but details were not presented.

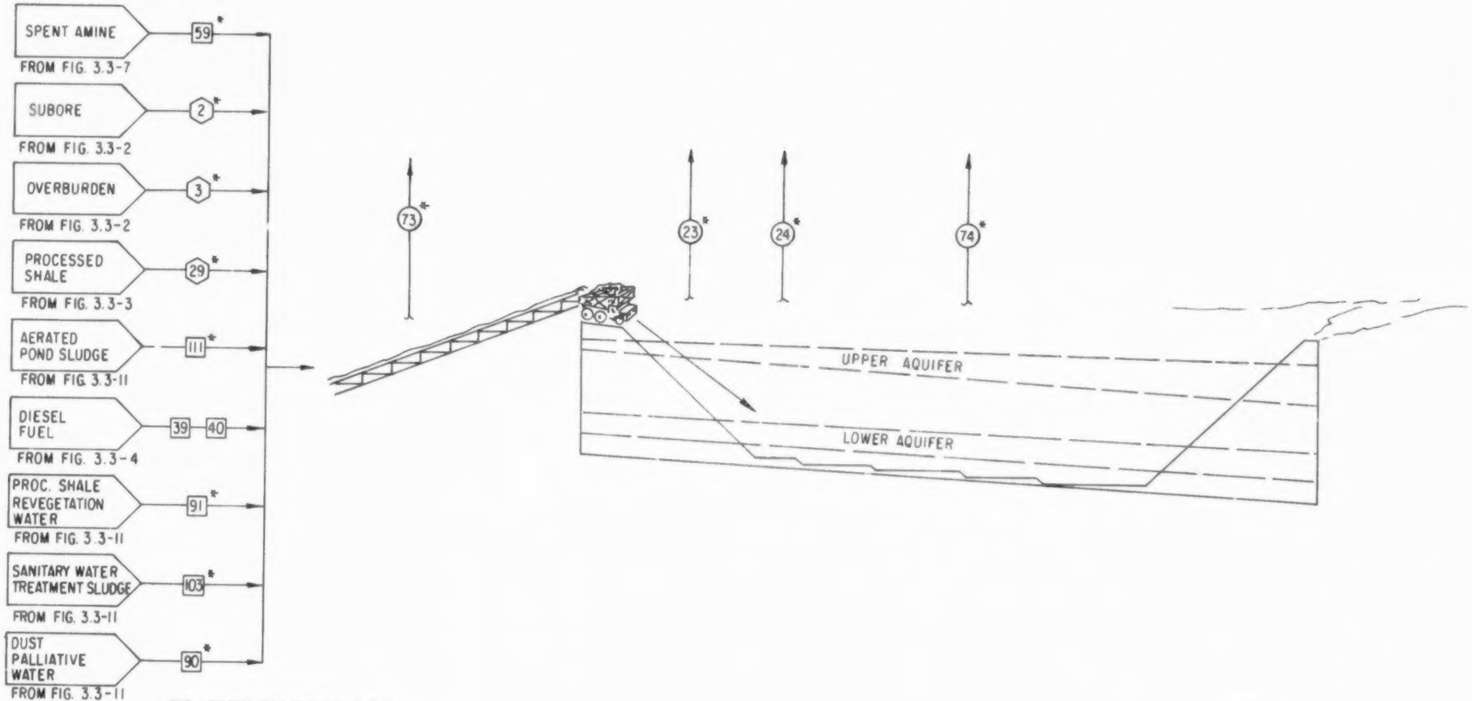
The subgrade ore, overburden, and processed shale (streams 2, 3 and 29) constitute the majority of the wastes. Several wastewaters, such as stripped liquor (stream 70), cooling tower blowdown (stream 105), boiler blowdown (stream 28), boiler feedwater treatment concentrate (stream 104), mine water clarifier sludge (stream 96), storm runoff (stream 93), and service and fire water (stream 95), are used to moisturize the processed shale to a 19% moisture content before disposal. The water management diagrams (see Figures 3.3-11 and 3.3-12) indicate the makeup of the moisturizing water (stream 78).

The waste transfer to the pit will be carried out in two phases. First, the waste material from the processing facility will be transported to the site using covered conveyors. Then, it will be loaded into 150-ton trucks and hauled into the pit. The backfilling operation will begin at the back of the pit, away from the mining operation. The pile will be constructed in 25-foot benches at 50-foot vertical intervals, using a slope of 2:1 (2 units horizontal: 1 unit vertical). The runoff will be collected (during the backfilling operation only) using runoff collection sumps located at the junction of the pile and the pit walls. As the pile reaches surface level, revegetation of the area will be carried out.

Transport of the processed shale to the pit will generate some particulate emissions at the conveyor transfer points (stream 73) and at the load-out hoppers (stream 23). These emissions are controlled with baghouses. The backfilling operation will also create fugitive dust (stream 74) which is controlled by the use of dust palliatives (stream 90). The diesel fuel used to operate the disposal equipment will create diesel emissions (stream 24).

3.3.10 Water Management

The water management plan for the Lurgi-Open Pit plant complex is presented in Figures 3.3-11 and 3.3-12. Groundwater (stream 4) collected from the mine dewatering operation is clarified prior to use or further treatment. A portion of the clarified mine water is used as sanitary/potable water (stream 99), fire and service water (stream 95), process makeup water (stream 87), etc., while the remainder is aerated and then discharged (stream 76). Before use in the cooling tower (stream 86), water is treated to retard biological growth and minimize scaling. Treated process waters from the plant, blowdowns, and concentrates are recycled to appropriate uses. The equalization pond serves as a source of water for processed shale moistening (stream 78). Sanitary wastes are treated by conventional biological processes; the water (stream 102) is then used for processed shale moisturizing and the sludge (stream 103) is used in revegetation.

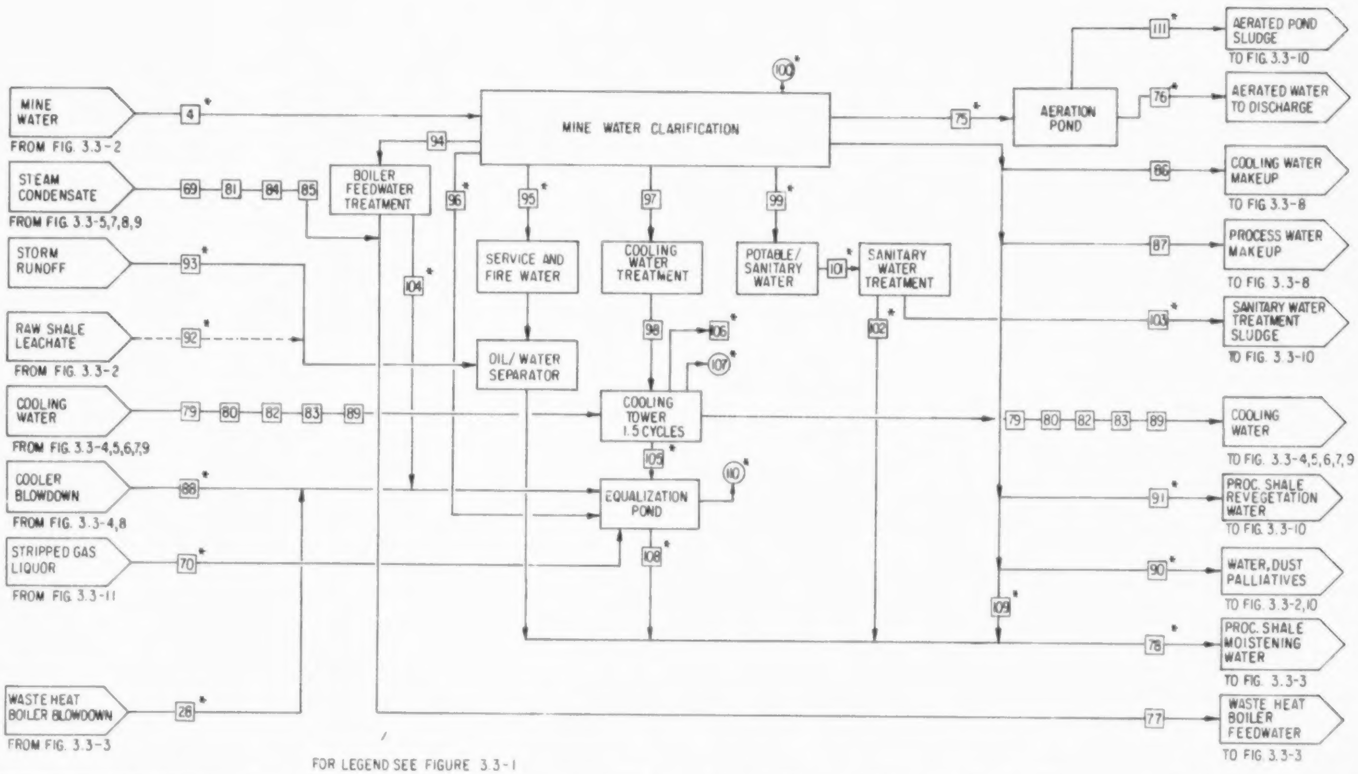

SPENT AMINE
FROM FIG. 3.3-7
59*
SUBORE
FROM FIG. 3.3-2
2*
OVERBURDEN
FROM FIG. 3.3-2
3*
PROCESSED SHALE
FROM FIG. 3.3-3
29*
AERATED POND SLUDGE
FROM FIG 3.3-11
111*
DIESEL FUEL
FROM FIG. 3.3-4
39
40
PROC. SHALE REVEGETATION WATER
FROM FIG. 3.3-11
91*
SANITARY WATER TREATMENT SLUDGE
FROM FIG. 3.3-11
103*
DUST PALLIATIVE WATER
FROM FIG. 3.3-11
90*
73*
23*
24*
74*
UPPER AQUIFER
LOWER AQUIFER

FOR LEGEND SEE FIGURE 3.3-1

STREAM NO.	2*	3*	23*	24*	29*	39	40	59*	73*	74*	90*	91*	103*	111*
STREAM IDENTITY	SUBORE	OVERBURDEN	BAGHOUSE EMISSION	DIESEL EMISSIONS	PROCESSED SHALE	DIESEL FUEL MINE EQUIP.	DIESEL FUEL DISP. EQUIP.	SPENT AMINE	BAGHOUSE EMISSION	FUGITIVE DUSTS	WATER, DUST PALLIATIVES	PROC. SHALE REVEGETATION WATER	SANIT. WATER TREATMENT SLUDGE	AERATED POND SLUDGE
FLOW RATE: 10^3ACFM 10^3lb/hr gpm	992	5175	64.5 8.3 lb/hr	151.3	9733	N.D.	N.D.	N.D.	64.6 0.3 lb/hr	52.0 lb/hr	1568	649	N.D.	N.D
TEMP, °F	AMB	AMB	AMB	N.D.	180	AMB	AMB	N.D.	AMB	←			→	AMB
PRESS., psig	AMB	←											→	AMB AMB

SOURCE: DRI based on Gulf Oil Corp. and
Standard Oil Co. (Indiana), March 1976

FIGURE 3.3-10 SOLID WASTE DISPOSAL


MINE WATER
FROM FIG. 3.3-2
STEAM CONDENSATE
FROM FIG. 3.3-5,7,8,9
STORM RUNOFF
RAW SHALE LEACHATE
FROM FIG. 3.3-2
COOLING WATER
FROM FIG. 3.3-4,5,6,7,9
COOLER BLOWDOWN
FROM FIG. 3.3-4,8
STRIPPED GAS LIQUOR
FROM FIG. 3.3-11
WASTE HEAT BOILER BLOWDOWN
FROM FIG. 3.3-3
4*
69
81
84
85
93*
92*
79
80
82
83
89
88*
70*
28*
94
BOILER FEEDWATER TREATMENT
104*
96*
MINE WATER CLARIFICATION
100*
95*
97
99*
SERVICE AND FIRE WATER
COOLING WATER TREATMENT
POTABLE/ SANITARY WATER
101*
SANITARY WATER TREATMENT
102*
98
106*
107*
OIL/ WATER SEPARATOR
COOLING TOWER 1.5 CYCLES
105*
110*
EQUALIZATION POND
108*
75*
AERATION POND
111*
76*
86
87
103*
109*
91*
90*
78*
77
AERATED POND SLUDGE
TO FIG. 3.3-10
AERATED WATER TO DISCHARGE
COOLING WATER MAKEUP
TO FIG. 3.3-8
PROCESS WATER MAKEUP
TO FIG. 3.3-8
SANITARY WATER TREATMENT SLUDGE
TO FIG. 3.3-10
COOLING WATER
TO FIG. 3.3-4,5,6,7,9
PROC. SHALE REVEGETATION WATER
TO FIG. 3.3-10
WATER, DUST PALLIATIVES
TO FIG. 3.3-2,10
PROC. SHALE MOISTENING WATER
TO FIG. 3.3-3
WASTE HEAT BOILER FEEDWATER
TO FIG. 3.3-3

FOR LEGEND SEE FIGURE 3.3-1

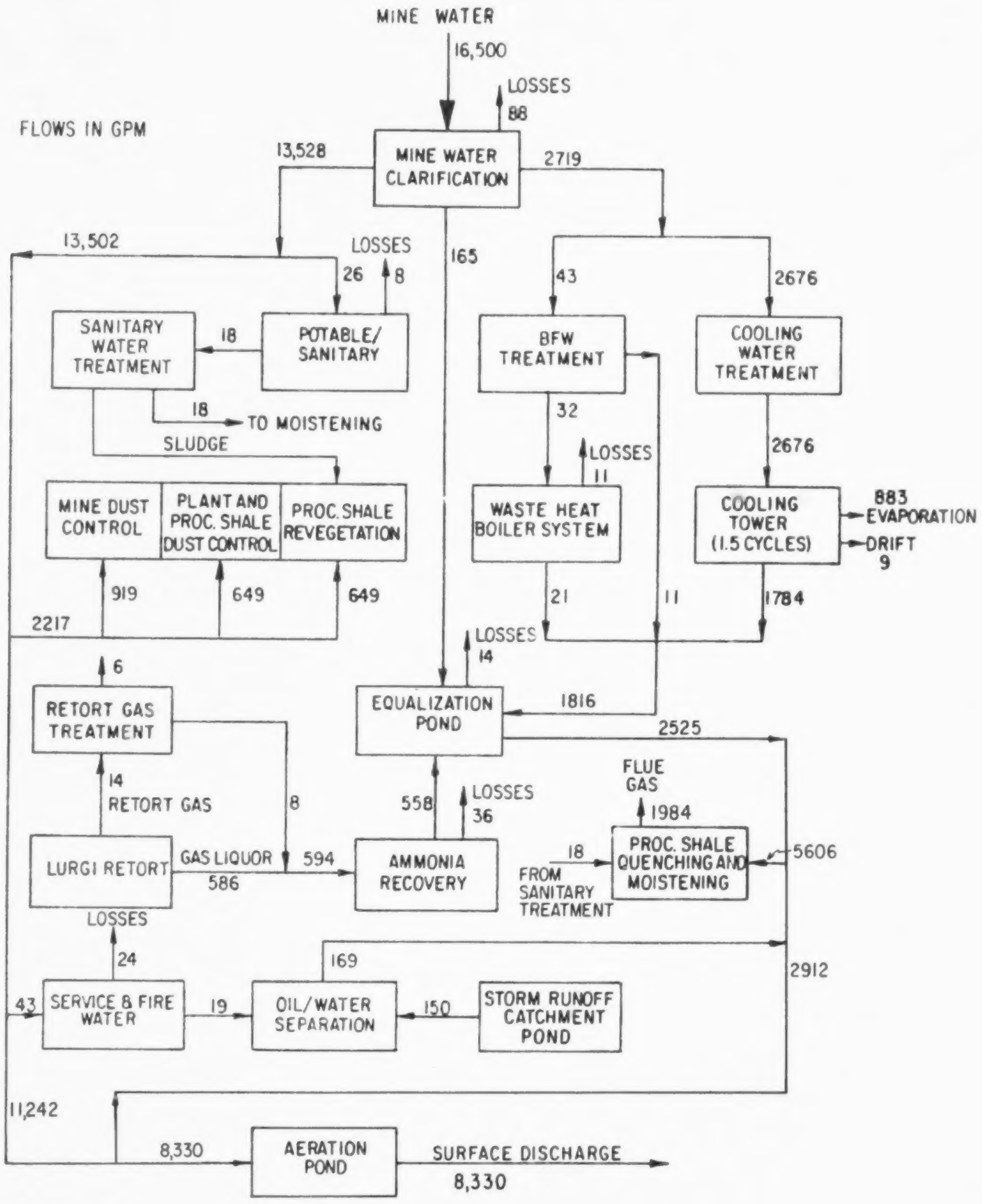
STREAM NO	4*	28*	70*	75*	76*	77	78*	79	80	82	83	89	69	81	84	85
STREAM IDENTITY	MINE WATER	WASTE HEAT BOILER BLOWDN	STRIP GAS LIQUOR	EXCES. MINE WATER	AERATED WATER TO DISCHRG	WASTE HEAT BOILR FD.WTR	PROC. SHALE MOIST. WATR	COOLING WATER					STEAM CONDENSATE			
FLOW RATE 10^3 ACFM																
10^3 lb/hr			2797													
gpm	16,500	21	558	8330	8330	2120	5624	325	5	8	66	1080	106	20	260	2
TEMP, °F	AMB	N.D	180	AMB	AMB	AMB	AMB	AMB					N.D			
PRESS., psig	AMB	AMB	6	AMB	←											

STREAM NO	86	87	88*	90*	91*	92*	93*	94	95*	96*	97	98	99*
STREAM IDENTITY	COOL. WTR MKUP.	PROC WTR MKUP.	COOL. BLW DN	WTR DUST PALL.	PROC SHLE. REVEG. WTR.	RAW SHLE. LEAC.	STRM RUN-OFF	BOIL FEED-WATR MKUP	SRVC. & FIRE WATR.	MINE WATR CLAR SLDG	WTR TO COOL TWR MAKEUP TREAT	TREAT WATR TO COOL TOWER	POTAB/ SANIT WATR
FLOW RATE 10^3 ACFM													
10^3 lb/hr													
gpm	2	3	661	1568	649	N.D.	150	43	43	165	2676	2676	26
TEMP, °F	AMB	AMB	N.D	AMB	←								
PRESS., psig													

STREAM NO	(100)*	101*	102*	103*	104*	105*	106*	(107)*	108*	109*	(110)*	111*
STREAM IDENTITY	WATR. EVAP. FROM CLARI.	USED SANIT. WATR.	TREAT SANIT. WATR.	SANIT. WATR TREAT SLDG	BOIL. FD.WTR TREAT CONC.	COOL. TWR BLW-DN	COOL. TWR DRFT	COOL TWR EVAP	EQUA. POND DISCH.	CLARI. WATR TO SHL. MOIST.	WTR. EVAP.	AERTD POND SLDG
FLOW RATE 10^3 ACFM												
10^3 lb/hr	44							442			7	N.D.
gpm		18	18	N.D.	11	1123	9		2525	2912		
TEMP, °F												AMB
PRESS., psig												AMB

SOURCE: DRI based on information provided by WPA

FIGURE 3.3-11 WATER MANAGEMENT


MINE WATER
16,500
LOSSES
88
FLOWS IN GPM
13,528
MINE WATER CLARIFICATION
2719
13,502
LOSSES
26
8
165
43
2676
SANITARY WATER TREATMENT
18
POTABLE/ SANITARY
BFW TREATMENT
COOLING WATER TREATMENT
18
TO MOISTENING
32
LOSSES
11
2676
SLUDGE
MINE DUST CONTROL
PLANT AND PROC. SHALE DUST CONTROL
PROC. SHALE REVEGETATION
WASTE HEAT BOILER SYSTEM
COOLING TOWER (1.5 CYCLES)
883
EVAPORATION
DRIFT
9
919
649
649
21
11
1784
2217
LOSSES
14
6
RETORT GAS TREATMENT
EQUALIZATION POND
1816
2525
14
RETORT GAS
8
558
LOSSES
36
FLUE GAS
1984
LURGI RETORT
GAS LIQUOR
586
594
AMMONIA RECOVERY
18
FROM SANITARY TREATMENT
PROC. SHALE QUENCHING AND MOISTENING
5606
LOSSES
24
169
2912
43
SERVICE & FIRE WATER
19
OIL/WATER SEPARATION
150
STORM RUNOFF CATCHMENT POND
11,242
8,330
AERATION POND
SURFACE DISCHARGE
8,330

SOURCE: WPA

FIGURE 3.3-12 OVERALL WATER MANAGEMENT SCHEME

SECTION 4

INVENTORY AND COMPOSITION OF PLANT PROCESS AND WASTE STREAMS

The stream compositions presented in this section were derived, to the extent possible, from pilot plant test data. In the absence of data from actual source testing, engineering analyses (by Denver Research Institute, Stone and Webster Engineering Corporation and Water Purification Associates) were performed on the technology and raw stream information from proposed industrial developments. The sources of these data, whether actual, estimated, or derived from published or unpublished information, are indicated.

The data presented are internally consistent for the overall plant complex; i.e., the principal chemical elements involved in emissions, effluents, and wastes are balanced throughout the plant. Trace elements generally are not considered because of the lack of consistent data available as a starting point. The stream compositions derived by engineering analysis generally agree with the available data from published sources. *Therefore, the data presented in this section, even though partly derived by engineering analysis, are believed to be both representative of the actual operations of such a plant and accurate enough to lead to relevant conclusions in analyses of various pollution controls.*

4.1 INVENTORY OF STREAMS

All but the most minor streams in the plant complex are inventoried in this section, and quantitative data are presented to define important characteristics of the streams. Section 4.2 presents detailed compositions of the major streams and shows changes in composition, from one point to the next, throughout the plant.

The streams encountered during the analysis of pollution control technologies for the plant are listed, along with their flow rates and components of concern, in Tables 4.1-1 (gases), 4.1-3 (liquids) and 4.1-5 (solids). Whether or not a stream must be controlled will depend upon its size, the quantities and characteristics of components, their allowable limits if released into the environment, and the disposition of the stream in an integrated plant design.

Tables 4.1-2, 4.1-4, and 4.1-6 list the major constituents in the streams. The streams are likewise divided into gases, liquids, and solids based on their physical characteristics. These tables summarize the data presented in Section 4.2, allowing for a quick comparison of the streams.

TABLE 4.1-1. INVENTORY OF GASEOUS STREAMS

Stream Number (Table No.)	Description of Stream	Mass Flow, 10^3 lb/hr (10^3 ACFM)	Components of Concern	Mass Flow of Component, lb/hr	Remarks[a]
5*	Primary Crusher (ore), Baghouse Emission	(122.2)	Particulates	15.7	Baghouse controlled.
6*	Primary Crusher (subore), Baghouse Emission	(12.2)	Particulates	1.6	Baghouse controlled.
7*	Primary Crusher (overburden), Baghouse Emission	(63.8)	Particulates	8.2	Baghouse controlled.
8*	Raw Shale Conveyor Transfer Point, Baghouse Emission	(72.6)	Particulates	0.4	Baghouse controlled.
9*	Swinging Boom Stacker, Baghouse Emission	(121.5)	Particulates	0.6	Baghouse controlled.
10*	Coarse Ore Conveyor Transfer Point, Baghouse Emission	(40.4)	Particulates	0.2	Baghouse controlled.
11*	Secondary Crusher, Baghouse Emission	(558.4)	Particulates	71.9	Baghouse controlled.
12*	Secondary Crushing to Screening Conveyor Transfer Point, Baghouse Emission	(40.4)	Particulates	0.2	Baghouse controlled.
13*	Secondary Screening, Baghouse Emission	(558.4)	Particulates	71.9	Baghouse controlled.
14*	Secondary Screening Conveyor Transfer Point, Baghouse Emission	(40.4)	Particulates	0.2	Baghouse controlled.
15*	Tertiary Crusher, Baghouse Emission	(628.2)	Particulates	80.8	Baghouse controlled.
16*	Tertiary Crushing to Tertiary Screening Conveyor Transfer Point, Baghouse Emission	(80.8)	Particulates	0.4	Baghouse controlled.
17*	Tertiary Screening, Baghouse Emission	(628.2)	Particulates	80.8	Baghouse controlled.
18*	Tertiary Screening to Fine Ore Storage Conveyor Transfer Point, Baghouse Emission	(40.4)	Particulates	0.2	Baghouse controlled.
19*	Fine Ore Storage, Baghouse Emission	(28.8)	Particulates	3.7	Baghouse controlled.

(Continued)

TABLE 4.1-1 (cont.)

Stream Number (Table No.)	Description of Stream	Mass Flow, 10^3 lb/hr (10^3 ACFM)	Components of Concern	Mass Flow of Component, lb/hr	Remarks[a]
20*	Retort Feed Hopper Conveyor Transfer Point, Baghouse Emission	(40.4)	Particulates	0.2	Baghouse controlled.
21*	Retort Feed Hopper, Baghouse Emission	(212.4)	Particulates	27.3	Baghouse controlled.
23*	Processed Shale Load-out Hopper, Baghouse Emission	(64.5)	Particulates	8.3	Baghouse controlled.
24*	Diesel Emissions	(151.3)	CO NOx SO_2 Hydrocarbons Particulates	34.8 469.9 35.6 12.7 33.0	Catalytic converters are installed on the diesel-operated equipment.
25	Combustion Air - Lift Pipes	5,439	--[b]	--	The air is preheated in the waste heat boiler and then used in the lift pipes for processed shale incineration.
26 (4.2-9)	Retort Vapors	1,425.2	Processed Shale Dust NH_3 H_2S SO_2 COS	78,600 1,924 1,197 256 N.D.	Oil products are dedusted before blending. Ammonia and sulfur dioxide are removed by subsequent gas liquor condensation.
27	High Pressure Steam	1,060	--	--	The steam is raised from the treated mine water and should be pure. Approximately 194 × 10^3 lb/hr of the steam are used for generating electrical power.
30	Steam to Humidifier	992	N.D.[c]	N.D.	The steam is produced during processed shale quenching and it may contain some entrained dust. It is added to the flue gas.
31* (4.2-19)	Lurgi Flue Gas	7,202.6	CO NOx SO_2 Hydrocarbons Particulates	657 2,440 500 6,262 1,107	Approximately 93% of the SO_2 is irreversibly adsorbed on the processed shale. The particulates are controlled by an electrostatic precipitator. RBOSC's modified DDP indicates emission of a large amount of hydrocarbons, measured as methane.

(Continued)

TABLE 4.1-1 (cont.)

Stream Number (Table No.)	Description of Stream	Mass Flow, 10^3 lb/hr (10^3 ACFM)	Components of Concern	Mass Flow of Component, lb/hr	Remarks[a]
32*	Raw Shale Retort Feed Conveyor, Baghouse Emission	(239.2)	Particulates	1.2	Baghouse controlled.
33 (4.2-11)	Raw Retort Gas	738.8	NH_3 H_2S SO_2	1,924 1,197 256	Ammonia and sulfur dioxide are removed from the retort gas by subsequent water scrubbing.
34 (4.2-12)	Retort Gas	149.0	NH_3 H_2S SO_2	29 697 19	Condensation of water vapor along with light oils results in removal of substantial quantities of ammonia and sulfur dioxide. The mass flows given are for the net retort gas after subtracting the amount sent to the lift pipes as supplemental fuel.
35	Retort Gas to Lift Pipes	107.0	NH_3 H_2S SO_2	21 501 14	The gas is supplied to the lift pipes as auxiliary fuel to support combustion. NOx and SO_2 emissions from the lift pipes may be somewhat increased, but SO_2 is controlled to 93% by adsorption on the processed shale.
44*	Fugitive Hydrocarbon Emissions from Storage Tanks	N.D.	Hydrocarbons	65.5	Proper storage tanks are used to prevent excessive hydrocarbon emissions. Includes stream 47.
45 (4.2-13)	Naphtha-free Retort Gas	121.7	NH_3 H_2S SO_2 CO_2 H_2O	29 697 19 55,981 3,832	The acid gases and moisture must be removed to achieve pipeline quality for the retort gas.
47*	Hydrocarbon Emissions from Naphtha Storage	N.D.	Hydrocarbons	N.D.	Included in stream 44.
48 (4.2-15)	Compressed Naphtha-free Gas	118.0	NH_3 H_2S CO_2 H_2O	2 697 55,964 97	Ammonia and water are reduced significantly by compression and cooling. Further drying is still necessary. Hydrogen sulfide and carbon dioxide also must be removed.
50	Steam to Naphtha Recovery	10	--	--	The steam is produced from the softened and demineralized mine water.

(Continued)

TABLE 4.1-1 (cont.)

Stream Number (Table No.)	Description of Stream	Mass Flow, 10^3 lb/hr (10^3 ACFM)	Components of Concern	Mass Flow of Component, lb/hr	Remarks[a]
51	Steam to DEA Unit	130	--	--	See stream 50.
52	Steam to Stretford Unit	1	--	--	See stream 50.
53	Steam to Ammonia Recovery Unit	53	--	--	See stream 50.
56 (4.2-16)	Sweet Gas from DEA Unit	61.9	NH_3 H_2S CO_2 H_2O	2 0.3 620 86	A majority of carbon dioxide and hydrogen sulfide is removed by absorption in DEA. The treated gas should be dried before pipelining.
57 (4.2-17)	Dried Fuel Gas to Pipeline	61.8	NH_3 H_2S CO_2 H_2O	2 0.3 620 9	Triethylene glycol (TEG) is used for absorbing most of the moisture in the gas. The dry gas may contain a small amount of TEG.
58 (4.2-18)	Acid Gases from DEA Regeneration	57.1	H_2S	696	The acid gases are treated by the Stretford process before being released to the atmosphere.
60*	TEG Regeneration Vent Emission	N.D.	N.D.	N.D.	The emission is water vapor with a small amount of TEG.
61 (4.2-18)	Stripping Air to Stretford	20.9	--	--	
63* (4.2-18)	Stretford Treated Acid Gases	53.0	H_2S	1.3	A majority of the hydrogen sulfide has been removed from the acid gases. The amount of H_2S emitted is negligible.
64 (4.2-18)	Stretford Oxidizer Vent Gas	24.3	--	--	The vent gas is primarily air, with some carbon dioxide and water vapor. It is used as a source of combustion air for the lift pipes.
72 (4.2-21)	Ammonia Overhead Vapors	20.2	NH_3 Organics	14 66	The overhead vapors are added to the lift pipes for the combustion of organics. As a result, the NOx emission from the lift pipes may be slightly increased.
73*	Processed Shale Conveyor Transfer Point, Baghouse Emission	(64.6)	Particulates	0.3	Baghouse controlled.

(Continued)

TABLE 4.1-1 (cont.)

Stream Number (Table No.)	Description of Stream	Mass Flow, 10^3 lb/hr (10^3 ACFM)	Components of Concern	Mass Flow of Component, lb/hr	Remarks[a]
74*	Fugitive Dusts	19,650	Particulates	52.0	Fugitive dusts emanate from mining, hauling, open storage, disposal, etc. These are controlled by water and foam sprays.
100*	Water Evaporation from Mine Water Clarifier	44	--	--	The evaporation is essentially pure water vapor, as clarified mine water is used.
107*	Cooling Tower Evaporation	442	--	--	See stream 100.
110*	Water Evaporation from Equalization Pond	7	--	--	See stream 100.
112*	Miscellaneous HC Emission	N.D.	Hydrocarbons	36.5	This emission represents the leakage from valves, pumps, etc. Proper maintenance practices are used to control the leakage.

* Indicates streams that come into contact with the environment.

[a] The remarks indicate the stream disposition. The control technologies applied to the streams are those proposed for the Lurgi-Open Pit technologies.

[b] Dashes (--) indicate no known components of concern.

[c] N.D. = Not determined.

Source: DRI estimates based on information from Gulf Oil Corp. and Standard Oil Co. (Indiana), March 1976, and Rio Blanco Oil Shale Co., February 1981.

TABLE 4.1-2. COMPOSITIONS OF GASEOUS STREAMS

Stream Number (Table No.)	Stream Description	Mass Flow, 10^3 lb/hr (10^3 ACFM)	Components, 10^3 lb/hr											
			H_2	CO	CO_2	N_2	NH_3	H_2S	CH_4	C_2H_4	C_2H_6	C_3H_6	C_3H_8	C_4H_8
5*	Primary Crusher (ore), Baghouse Emission	(122.2)	0**	0	0	N.D.**	0	0	N.D.	0	0	0	0	0
6*	Primary Crusher (subore), Baghouse Emission	(12.2)	0	0	0	N.D.	0	0	N.D.	0	0	0	0	0
7*	Primary Crusher (overburden), Baghouse Emission	(63.8)	0	0	0	N.D.	0	0	N.D.	0	0	0	0	0
8*	Raw Shale Conveyor Transfer Point, Baghouse Emission	(72.6)	0	0	0	N.D.	0	0	N.D.	0	0	0	0	0
9*	Swinging Boom Stacker, Baghouse Emission	(121.5)	0	0	0	N.D.	0	0	N.D.	0	0	0	0	0
10*	Coarse Ore Conveyor Transfer Point, Baghouse Emission	(40.4)	0	0	0	N.D.	0	0	N.D.	0	0	0	0	0
11*	Secondary Crusher, Baghouse Emission	(558.4)	0	0	0	N.D.	0	0	N.D.	0	0	0	0	0
12*	Secondary Crushing to Screening Conveyor Transfer Point, Baghouse Emission	(40.4)	0	0	0	N.D.	0	0	N.D.	0	0	0	0	0
13*	Secondary Screening, Baghouse Emission	(558.4)	0	0	0	N.D.	0	0	N.D.	0	0	0	0	0
14*	Secondary Screening Conveyor Transfer Point, Baghouse Emission	(40.4)	0	0	0	N.D.	0	0	N.D.	0	0	0	0	0
15*	Tertiary Crusher, Baghouse Emission	(628.2)	0	0	0	N.D.	0	0	N.D.	0	0	0	0	0
16*	Tertiary Crushing to Tertiary Screening Conveyor Transfer Point, Baghouse Emission	(80.8)	0	0	0	N.D.	0	0	N.D.	0	0	0	0	0
17*	Tertiary Screening, Baghouse Emission	(628.2)	0	0	0	N.D.	0	0	N.D.	0	0	0	0	0
18*	Tertiary Screening to Fine Ore Storage Conveyor Transfer Point, Baghouse Emission	(40.4)	0	0	0	N.D.	0	0	N.D.	0	0	0	0	0

(Continued)

TABLE 4.1-2 (cont.)

Stream Number (Table No.)	Stream Description	Mass Flow, 10^3 lb/hr (10^3 ACFM)	Components, 10^3 lb/hr													
			C_4H_{10}	Misc. HC	C_4^+	Light Oil	Middle Oil	Heavy Oil	COS	CS_2	CH_3SH	H_2O	O_2	NOx lb/hr	SO_2 lb/hr	TPM lb/hr
5*	Primary Crusher (ore), Baghouse Emission	(122.2)	0	0	0	0	0	0	0	0	0	N.D.	N.D.	0	0	15.7
6*	Primary Crusher (subore), Baghouse Emission	(12.2)	0	0	0	0	0	0	0	0	0	N.D.	N.D.	0	0	1.6
7*	Primary Crusher (overburden), Baghouse Emission	(63.8)	0	0	0	0	0	0	0	0	0	N.D.	N.D.	0	0	8.2
8*	Raw Shale Conveyor Transfer Point, Baghouse Emission	(72.6)	0	0	0	0	0	0	0	0	0	N.D.	N.D.	0	0	0.4
9*	Swinging Boom Stacker, Baghouse Emission	(121.5)	0	0	0	0	0	0	0	0	0	N.D.	N.D.	0	0	0.6
10*	Coarse Ore Conveyor Transfer Point, Baghouse Emission	(40.4)	0	0	0	0	0	0	0	0	0	N.D.	N.D.	0	0	0.2
11*	Secondary Crusher, Baghouse Emission	(558.4)	0	0	0	0	0	0	0	0	0	N.D.	N.D.	0	0	71.9
12*	Secondary Crushing to Screening Conveyor Transfer Point, Baghouse Emission	(40.4)	0	0	0	0	0	0	0	0	0	N.D.	N.D.	0	0	0.2
13*	Secondary Screening, Baghouse Emission	(558.4)	0	0	0	0	0	0	0	0	0	N.D.	N.D.	0	0	71.9
14*	Secondary Screening Conveyor Transfer Point, Baghouse Emission	(40.4)	0	0	0	0	0	0	0	0	0	N.D.	N.D.	0	0	0.2
15*	Tertiary Crusher, Baghouse Emission	(628.2)	0	0	0	0	0	0	0	0	0	N.D.	N.D.	0	0	80.8
16*	Tertiary Crushing to Tertiary Screening Conveyor Transfer Point, Baghouse Emission	(80.8)	0	0	0	0	0	0	0	0	0	N.D.	N.D.	0	0	0.4
17*	Tertiary Screening, Baghouse Emission	(628.2)	0	0	0	0	0	0	0	0	0	N.D.	N.D.	0	0	80.8
18*	Tertiary Screening to Fine Ore Storage Conveyor Transfer Point, Baghouse Emission	(40.4)	0	0	0	0	0	0	0	0	0	N.D.	N.D.	0	0	0.2

(Continued)

TABLE 4.1-2 (cont.)

Stream Number (Table No.)	Stream Description	Mass Flow, 10^3 lb/hr (10^3 ACFM)	Components, 10^3 lb/hr											
			H_2	CO	CO_2	N_2	NH_3	H_2S	CH_4	C_2H_4	C_2H_6	C_3H_6	C_3H_8	C_4H_8
19*	Fine Ore Storage, Baghouse Emission	(28.8)	0	0	0	N.D.	0	0	N.D.	0	0	0	0	0
20*	Retort Feed Hopper Conveyor Transfer Point, Baghouse Emission	(40.4)	0	0	0	N.D.	0	0	N.D.	0	0	0	0	0
21*	Retort Feed Hopper, Baghouse Emission	(212.4)	0	0	0	N.D.	0	0	N.D.	0	0	0	0	0
23*	Processed Shale Load-out Hopper, Baghouse Emission	(64.5)	0	0	0	N.D.	0	0	N.D.	0	0	0	0	0
24*	Diesel Emissions	(151.3)	0	34.8 lb/hr	N.D.	N.D.	0	0	0	0	0	0	0	0
25	Combustion Air - Lift Pipes	5,439	0	0	0	4,146	0	0	0	0	0	0	0	0
26 (4.2-9)	Retort Vapors	1,425.2	4.97	7.14	98.46	7.14	1.92	1.20	18.95	12.24	12.57	15.68	8.42	13.77
27	High Pressure Steam	1,060	0	0	0	0	0	0	0	0	0	0	0	0
30	Steam to Humidifier	992	0	0	0	0	0	0	0	0	0	0	0	0
31* (4.2-19)	Lurgi Flue Gas	7,206.6	0	0.66	1,443.63	4,170.9	0	0	0	0	0	0	0	0
32*	Raw Shale Retort Feed Conveyor, Baghouse Emission	(239.2)	0	0	0	N.D.	0	0	N.D.	0	0	0	0	0
33 (4.2-11)	Raw Retort Gas	738.8	4.97	7.14	98.46	7.14	1.92	1.20	18.95	12.24	12.57	15.68	8.42	13.77
34 (4.2-12)	Retort Gas	149.0	2.89	4.16	55.98	4.16	0.03	0.70	11.03	7.13	7.32	9.13	4.90	8.01
35	Retort Gas to Lift Pipes	107.0	2.08	2.99	40.2	2.99	0.02	0.50	7.92	5.12	5.26	6.56	3.52	5.75
44*	Fugitive Hydrocarbon Emissions from Storage Tanks (stream 47 included)	N.D.	0	0	0	0	0	0	0	0	0	0	0	0
45 (4.2-13)	Naphtha-free Retort Gas	121.7	2.89	4.16	55.98	4.16	0.03	0.70	11.03	7.13	7.32	9.13	4.90	8.01

(Continued)

TABLE 4.1-2 (cont.)

Stream Number (Table No.)	Stream Description	Mass Flow, 10^3 lb/hr (10^3 ACFM)	Components, 10^3 lb/hr C_4H_{10}	Misc. HC	C_4^+	Light Oil	Middle Oil	Heavy Oil	COS	CS_2	CH_3SH	H_2O	O_2	NOx lb/hr	SO_2 lb/hr	TPM lb/hr
19*	Fine Ore Storage, Baghouse Emission	(28.8)	0	0	0	0	0	0	0	0	0	N.D.	N.D.	0	0	3.7
20*	Retort Feed Hopper Conveyor Transfer Point, Baghouse Emission	(40.4)	0	0	0	0	0	0	0	0	0	N.D.	N.D.	0	0	0.2
21*	Retort Feed Hopper, Baghouse Emission	(212.4)	0	0	0	0	0	0	0	0	0	N.D.	N.D.	0	0	27.[illegible]
23*	Processed Shale Load-out Hopper, Baghouse Emission	(64.5)	0	0	0	0	0	0	0	0	0	N.D.	N.D.	0	0	8.3
24*	Diesel Emissions	(151.3)	0	0.013	0	0	0	0	0	0	0	N.D.	N.D.	469.9	35.6	33
25	Combustion Air - Lift Pipes	5,439	0	0	0	0	0	0	0	0	0	35	1,258	0	0	0
26 (4.2-9)	Retort Vapors	1,425.2	4.23	0.24	23.87	205.77	413.24	196.51	N.D.	0	0	300.05	0	0	256	78,600
27	High Pressure Steam	1,060	0	0	0	0	0	0	0	0	0	1,060	0	0	0	0
30	Steam to Humidifier	992	0	0	0	0	0	0	0	0	0	992	0	0	0	N.D.
31* (4.2-19)	Lurgi Flue Gas	7,202.6	0	6.26	0	0	0	0	0	0	0	1,224	353.12	2,440	500	1,107
32*	Raw Shale Retort Feed Conveyor, Baghouse Emission	(239.2)	0	0	0	0	0	0	0	0	0	N.D.	N.D.	0	0	1.2
33 (4.2-11)	Raw Retort Gas	738.8	4.23	0.24	23.87	205.77	1.91	0	0	0	0	300.05	0	0	256	0
34 (4.2-12)	Retort Gas	149.0	2.46	13.10	13.89	0	0	0	0	0	0	4.10	0	0	19	0
35	Retort Gas to Lift Pipes	107.0	1.77	9.41	9.97	0	0	0	0	0	0	2.94	0	0	14	0
44*	Fugitive Hydrocarbon Emissions from Storage Tanks (stream 47 included)	N.D.	0	0.066	0	0	0	0	0	0	0	0	0	0	0	0
45 (4.2-13)	Naphtha-free Retort Gas	121.7	2.46	0	0	0	0	0	0	0	0	3.83	0	0	19	0

(Continued)

TABLE 4.1-2 (cont.)

Stream Number (Table No.)	Stream Description	Mass Flow, 10^3 lb/hr (10^3 ACFM)	Components, 10^3 lb/hr											
			H_2	CO	CO_2	N_2	NH_3	H_2S	CH_4	C_2H_4	C_2H_6	C_3H_6	C_3H_8	C_4H_8
47*	Hydrocarbon Emissions from Naphtha Storage (included in stream 44)	N.D.	0	0	0	0	0	0	0	0	0	0	0	0
48 (4.2-15)	Compressed Naphtha-free Gas	118.0	2.89	4.16	55.96	4.16	0.002	0.70	11.03	7.13	7.32	9.13	4.90	8.01
50	Steam to Naphtha Recovery	10	0	0	0	0	0	0	0	0	0	0	0	0
51	Steam to DEA Unit	130	0	0	0	0	0	0	0	0	0	0	0	0
52	Steam to Stretford Unit	1	0	0	0	0	0	0	0	0	0	0	0	0
53	Steam to Ammonia Recovery Unit	53	0	0	0	0	0	0	0	0	0	0	0	0
56 (4.2-16)	Sweet Gas from DEA Unit	61.9	2.89	4.16	0.62	4.16	0.002	0.0003	11.03	7.13	7.32	9.13	4.90	8.01
57 (4.2-17)	Dried Fuel Gas to Pipeline	61.8	2.89	4.16	0.62	4.16	0.002	0.0003	11.03	7.13	7.32	9.13	4.90	8.01
58 (4.2-18)	Acid Gases from DEA Regeneration	57.1	0	0	55.34	N.D.	0	0.70	0	0	0	0	0	0
60*	TEG Regeneration Vent Emission	N.D.	0	0	0	0	0	0	0	0	0	0	0	0
61 (4.2-18)	Stripping Air to Stretford	20.9	0	0	0	16.12	0	0	0	0	0	0	0	0
63* (4.2-18)	Stretford Treated Acid Gases	53.0	0	0	52	0	0	0.0013	0	0	0	0	0	0
64 (4.2-18)	Stretford Oxidizer Vent Gas	24.3	0	0	3.32	16.12	0	0	0	0	0	0	0	0
72 (4.2-21)	Ammonia Overhead Vapors	20.2	0	0	2.30	0	0.014	0	0	0	0	0	0	0
73*	Processed Shale Conveyor Transfer Point, Baghouse Emission	(64.6)	0	0	0	0	0	0	0	0	0	0	0	0
74*	Fugitive Dusts	19,650	0	0	0	0	0	0	0	0	0	0	0	0

(Continued)

TABLE 4.1-2 (cont.)

			Components, 10^3 lb/hr													
Stream Number (Table No.)	Stream Description	Mass Flow, 10^3 lb/hr (10^3 ACFM)	C_4H_{10}	Misc. HC	C_4'	Light Oil	Middle Oil	Heavy Oil	COS	CS_2	CH_3SH	H_2O	O_2	NOx lb/hr	SO_2 lb/hr	TPM lb/hr
47*	Hydrocarbon Emissions from Naphtha Storage (included in stream 44)	N.D.	0	N.D.	0	0	0	0	0	0	0	0	0	0	0	0
48 (4.2-15)	Compressed Naphtha-free Gas	118	2.46	0	0	0	0	0	0	0	0	0.10	0	0	0	0
50	Steam to Naphtha Recovery	10	0	0	0	0	0	0	0	0	0	10	0	0	0	0
51	Steam to DEA Unit	130	0	0	0	0	0	0	0	0	0	130	0	0	0	0
52	Steam to Stretford Unit	1	0	0	0	0	0	0	0	0	0	1	0	0	0	0
53	Steam to Ammonia Recovery Unit	53	0	0	0	0	0	0	0	0	0	53	0	0	0	0
56 (4.2-16)	Sweet Gas from DEA Unit	62	2.46	0	0	0	0	0	0	0	0	0.09	0	0	0	0
57 (4.2-17)	Dried Fuel Gas to Pipeline	62	2.46	0	0	0	0	0	0	0	0	9 lb/hr	0	0	0	0
58 (4.2-18)	Acid Gases from DEA Regeneration	57	0	0	0	0	0	0	0	0	0	1.03	0	0	0	0
60*	TEG Regeneration Vent Emission	N.D.	0	0	0	0	0	0	0	0	0	N.D.	0	0	0	0
61 (4.2-18)	Stripping Air to Stretford	20.9	0	0	0	0	0	0	0	0	0	0.13	4.63	0	0	0
63* (4.2-18)	Stretford Treated Acid Gases	53	0	0	0	0	0	0	0	0	0	0.96	0	0	0	0
64 (4.2-18)	Stretford Oxidizer Vent Gas	24.3	0	0	0	0	0	0	0	0	0	0.60	4.29	0	0	0
72 (4.2-21)	Ammonia Overhead Vapors	20.2	0	0.07	0	0	0	0	0	0	0	17.79	0	0	0	0
73*	Processed Shale Conveyor Transfer Point, Baghouse Emission	(64.6)	0	0	0	0	0	0	0	0	0	N.D.	0	0	0	0.3
74*	Fugitive Dusts	19,650	0	0	0	0	0	0	0	0	0	N.D.	0	0	0	52

(Continued)

TABLE 4.1-2 (cont.)

Stream Number (Table No.)	Stream Description	Mass Flow, 10^3 lb/hr (10^3 ACFM)	Components, 10^3 lb/hr											
			H_2	CO	CO_2	N_2	NH_3	H_2S	CH_4	C_2H_4	C_2H_6	C_3H_6	C_3H_8	C_4H_8
100*	Water Evaporation from Mine Water Clarifier	44	0	0	0	0	0	0	0	0	0	0	0	0
107*	Cooling Tower Evaporation	442	0	0	0	0	0	0	0	0	0	0	0	0
110*	Water Evaporation from Equalization Pond	7	0	0	0	0	0	0	0	0	0	0	0	0
112*	Miscellaneous HC Emission	N.D.	0	0	0	0	0	0	0	0	0	0	0	0

(Continued)

TABLE 4.1-2 (cont.)

Stream Number (Table No.)	Stream Description	Mass Flow, 10^3 lb/hr (10^3 ACFM)	Components, 10^3 lb/hr													
			C_4H_{10}	Misc. HC	C_4^+	Light Oil	Middle Oil	Heavy Oil	COS	CS_2	CH_3SH	H_2O	O_2	NOx lb/hr	SO_2 lb/hr	TPM lb/hr
100*	Water Evaporation from Mine Water Clarifier	44	0	0	0	0	0	0	0	0	0	44	0	0	0	0
107*	Cooling Tower Evaporation	442	0	0	0	0	0	0	0	0	0	442	0	0	0	0
110*	Water Evaporation from Equalization Pond	7	0	0	0	0	0	0	0	0	0	7	0	0	0	0
112*	Miscellaneous HC Emission	N.D.	0	0.037	0	0	0	0	0	0	0	N.D.	0	0	0	0

* Indicates streams that come into contact with the environment.

**0 = Estimated to be insignificant; N.D. = Not determined.

Source: DRI estimates based on information from Gulf Oil Corp. and Standard Oil Co. (Indiana), March 1976, and Rio Blanco Oil Shale Co., February 1981.

TABLE 4.1-3. INVENTORY OF LIQUID STREAMS

Stream Number (Table No.)	Description of Stream	Mass Flow, gpm (10^3 lb/hr)	Components of Concern	Mass Flow of Component, lb/hr	Remarks[a]
4* (4.2-22)	Mine Water	16,500	TDS Boron Phenol	8,266 5 0.02	The water is clarified and properly treated before use in the plant.
28*	Blowdown - Waste Heat Boiler	21	--[b]	--	Clarified mine water is softened and demineralized before use in the boiler. The blowdown is used for processed shale moisturizing.
36 (4.2-10)	Light Oils to Storage	406 (185.2)	Organic -Nitrogen -Sulfur	 16,000 7,000	The composition of individual oil fractions is not known. The quantities of nitrogen and sulfur indicated are for combined heavy oils, middle oils and light oils. Treatment may be required if on-site use as a fuel or upgrading is considered.
37	Light Oil Makeup to Naphtha Recovery	N.D.[c]	N.D.	N.D.	This is an internal recycle stream. Treatment is not necessary.
38 (4.2-10)	Middle Oils to Storage	937 (411.3)	Organic -Nitrogen -Sulfur	 -- --	See stream 36.
39	Diesel Fuel - Mining Equipment	N.D.	--	--	Lurgi middle oils are used as the diesel fuel.
40	Diesel Fuel - Disposal Equipment	N.D.	--	--	See stream 39.
41 (4.2-20)	Gas Liquor	586 (297.7)	Free NH_3 Fixed NH_3 Fixed SO_2	1,758 117 221	The volatile components are removed by steam stripping in the ammonia recovery unit. The nonvolatile components are eventually mixed with the processed shale.
42 (4.2-10)	Heavy Oils to Storage	416 (136.5)	Organic -Nitrogen -Sulfur	 -- --	See stream 36.
46 (4.2-14)	Naphtha Product to Storage	(27.3)	--	--	The amount indicated is the net product after usage in the lift pipes. The naphtha is made up of C_5^+ hydrocarbons.
49 (4.2-20)	Compressor Condensate	8 (3.8)	Free NH_3 Fixed NH_3 Fixed SO_2	17 9 16	The remarks for gas liquor (stream 41) also apply. The condensate is added to the gas liquor for further treatment.

(Continued)

TABLE 4.1-3 (cont.)

Stream Number (Table No.)	Description of Stream	Mass Flow, gpm (10^3 lb/hr)	Components of Concern	Mass Flow of Component, lb/hr	Remarks[a]
54	Amine Makeup	N.D.	--	--	The diethanolamine is added to make up the reagent losses.
55	TEG Makeup	N.D.	--	--	A small amount of TEG is lost during the reagent regeneration and is made up with the fresh chemical.
59*	Spent Amine	N.D.	N.D.	N.D.	The amine spent during the reagent regeneration is removed periodically and sent for disposal.
62	Stretford Chemicals	(0.03)	--	--	The Holmes-Stretford mix and soda ash are added to make up the reagent losses.
65	Stretford Spent Liquor to Reclaim	N.D.	N.D.	N.D.	The liquor is shipped for off-site disposal or the useful chemicals may be reclaimed.
66 (4.2-18)	Liquid Sulfur Product to Storage	(0.70) 7.6 LTPSD	--	--	Stretford sulfur is reported to have +99.9% purity.
67	Phosphoric Acid	(0.01)	--	--	This is a reagent makeup to the ammonia recovery process.
69	Steam Condensate from Ammonia Recovery	106	--	--	Softened and demineralized mine water is used for raising the steam. The steam is condensed upon use and returned to the boilers.
70* (4.2-21)	Stripped Gas Liquor	558 (279.7)	TDS NH_3 Dissolved Organics	471 4 170	The free and fixed ammonia in the gas liquor are recovered in the ammonia plant. Stripped liquor is used for processed shale moisturizing.
71 (4.2-21)	Anhydrous Ammonia to Storage	(1.9) 22.6 TPSD	NH_3 vapors	N.D.	Refrigerated storage tanks are used to reduce the NH_3 emissions.
75* (4.2-23)	Excess Mine Water to Aeration Pond	8,330	TDS Boron Phenol	4,170 2.6 0.004	This is the excess mine water after process needs. It is aerated to reduce the organics content, then discharged on the surface.
76* (4.2-23)	Aerated Water to Discharge	8,330	TDS Boron Phenol COD	4,170 2.6 0.004 50	The COD is reduced by 25% due to aeration. The treated water is discharged on the surface.

(Continued)

TABLE 4.1-3 (cont.)

Stream Number (Table No.)	Description of Stream	Mass Flow, gpm (10^3 lb/hr)	Components of Concern	Mass Flow of Component, lb/hr	Remarks[a]
77	Feedwater to Waste Heat Boiler	2,120	--	--	See stream 28.
78*	Total Processed Shale Moistening Water	5,624	TDS	N.D.	Various wastewater streams are combined and used for processed shale quenching and moisturizing.
79	Cooling Water to Lurgi Oil Recovery	325	--	--	Treated mine water is used for plant cooling requirements.
80	Cooling Water to Naphtha Recovery	5	--	--	See stream 79. The quantity given is to make up the losses.
81	Steam Condensate from Naphtha Stripper	20	--	--	See stream 28.
82	Cooling Water to Compression Cooling	8	--	--	See stream 79. The quantity given is to make up the losses.
83	Cooling Water to DEA-TEG Treatment	66	--	--	See stream 79. The quantity given is to make up the losses.
84	Steam Condensate from DEA-TEG Treatment	260	--	--	See stream 28.
85	Steam Condensate from Stretford	2	--	--	See stream 28.
86	Cooling Water Makeup to Stretford	2	--	--	See stream 79. The quantity given is to make up the losses.
87	Process Water Makeup to Stretford	3	--	--	Process water of boiler feedwater quality is used.
88*	Humidified Air Cooler Blowdown	661	--	--	Treated mine water is used in the humidified air coolers in the oil recovery and Stretford processes. The blowdown is used in processed shale moisturizing.
89	Cooling water to Ammonia Recovery	1,080	--	--	See stream 79. The quantity given is for the cooling water circulated.
90*	Water for Dust Palliatives	1,568	--	--	Clarified mine water is used for the raw and processed shale dust control.
91*	Processed Shale Revegetation Water	649	--	--	Clarified mine water is used.

(Continued)

TABLE 4.1-3 (cont.)

Stream Number (Table No.)	Description of Stream	Mass Flow, gpm (10^3 lb/hr)	Components of Concern	Mass Flow of Component, lb/hr	Remarks[a]
92* (4.2-3, 4.2-4)	Raw Shale Leachate	N.D.	TDS DOC	3,490 mg/l 13 mg/l	Leachate data are derived from a Tract C-a shale lysimeter study.
93*	Storm Runoff	150	N.D.	N.D.	Storm runoff water is collected and used for processed shale moisturizing.
94	Boiler Feedwater Makeup	43	--	--	Softened and demineralized mine water is used to compensate steam and blowdown losses.
95*	Service and Fire Water	43	--	--	Clarified mine water is used.
96*	Mine Water Clarifier Sludge	165	N.D.	N.D.	Suspended solids and debris are collected during the mine water clarification and used for processed shale moisturizing.
97	Water to Cooling Tower Makeup Treatment	2,676	TDS	1,340	Clarified mine water is treated with H_2SO_4 to retard the biological growth in the water and then used for plant cooling.
98	Treated Water to Cooling Tower	2,676	TDS	1,340	Treated mine water is used to cool the cooling water return from the plant.
99*	Potable/Sanitary Water	26	--	--	Clarified mine water is treated and used for the sanitary needs.
101*	Used Sanitary Water to Municipal Treatment	18	N.D.	N.D.	Used sanitary water is sent to municipal treatment before disposal.
102*	Treated Sanitary Water	18	--	--	The sanitary water after municipal treatment is used for processed shale moisturizing.
103*	Sanitary Water Treatment Sludge	N.D.	--	--	The sludge from the sanitary water treatment is dewatered, then used as a fertilizer in revegetation.
104*	Boiler Feedwater Treatment Concentrate	11	--	--	Regenerated waste from zeolite softening and demineralization is used for processed shale moisturizing.
105*	Cooling Tower Blowdown	1,123	--	--	Treated mine water is used for plant cooling requirements. The quantity given does not include the humidified air cooler blowdown (stream 88). The total blowdown would be 1,784 gpm.

(Continued)

TABLE 4.1-3 (cont.)

Stream Number (Table No.)	Description of Stream	Mass Flow, gpm (10^3 lb/hr)	Components of Concern	Mass Flow of Component, lb/hr	Remarks[a]
106*	Cooling Tower Drift	9	--	--	Treated mine water is used in the cooling tower. The drift is essentially pure water.
108*	Equalization Pond Discharge to Processed Shale Moistening	2,525	N.D.	N.D.	Various wastewaters (e.g., sludges, concentrates, blowdowns) are combined and used for processed shale moisturizing.
109*	Clarified Mine Water to Processed Shale Moistening	2,912	TDS	1,096	Clarified mine water is used to fulfill the processed shale moisturizing needs.
111*	Aerated Pond Sludge	N.D.	N.D.	N.D.	The sludge may contain some bio-oxidized material and settled inorganic salts. It is sent for processed shale moisturizing.

* Indicates streams that come into contact with the environment.

[a] The remarks indicate the stream disposition. The controls and treatments applied to the streams are those proposed for the Lurgi-Open Pit technologies.

[b] Dashes (--) indicate no known components of concern.

[c] N.D = Not determined.

Source: DRI estimates based on information from Gulf Oil Corp. and Standard Oil Co. (Indiana), March 1976, and Rio Blanco Oil Shale Co., February 1981.

TABLE 4.1-4. COMPOSITIONS OF LIQUID STREAMS

Stream Number (Table No.)	Stream Description	Mass Flow, gpm (10^3 lb/hr)	Components, lb/hr					
			CO_2	NH_3	H_2S	TDS & TSS	Organics	H_2O
4* (4.2-22)	Mine Water	16,500	N.D.**	0**	0	8,266	N.D.	8,250,000
28*	Blowdown - Waste Heat Boiler	21	0	0	0	22	N.D.	10,500
36 (4.2-10)	Light Oils to Storage	406 (185.2)	0	0	0	0	185,160	0
37	Light Oil Makeup to Naphtha Recovery	N.D.	0	0	0	0	N.D.	0
38 (4.2-10)	Middle Oils to Storage	937 (411.3)	0	0	0	0	411,330	0
39	Diesel Fuel - Mining Equipment	N.D.	0	0	0	0	N.D.	0
40	Diesel Fuel - Disposal Equipment	N.D.	0	0	0	0	N.D.	0
41 (4.2-20)	Gas Liquor	586 (297.7)	2,275	1,758	0	N.D.	236	293,000
42 (4.2-10)	Heavy Oils to Storage	416 (196.5)	0	0	0	0	196,510	0
46 (4.2-14)	Naphtha Product to Storage	(27.3)	0	0	0	0	26,990	270
49 (4.2-20)	Compressor Condensate	8 (3.8)	20	17	0	29	N.D.	3,735
54	Amine Makeup	N.D.	0	0	0	0	N.D.	N.D.
55	TEG Makeup	N.D.	0	0	0	0	N.D.	N.D.
59*	Spent Amine	N.D.	0	0	0	0	N.D.	N.D.
62	Stretford Chemicals	(0.03)	0	0	0	2.6	N.D.	N.D.
65	Stretford Spent Liquor to Reclaim	N.D.	0	0	N.D.	N.D.	N.D.	N.D.
66 (4.2-18)	Liquid Sulfur Product to Storage	(0.70) 7.6 LTPSD	0	0	0	0	0	0
67	Phosphoric Acid	(0.01)	0	0	0	0	0	N.D.
69	Steam Condensate from Ammonia Recovery	106	0	0	0	0	0	53,000
70* (4.2-21)	Stripped Gas Liquor	558 (279.7)	0	4	0	471	170	279,076

(Continued)

TABLE 4.1-4 (cont.)

Stream Number (Table No.)	Stream Description	Mass Flow, gpm (10^3 lb/hr)	Components, lb/hr					
			CO_2	NH_3	H_2S	TDS & TSS	Organics	H_2O
71 (4.2-21)	Anhydrous Ammonia to Storage	(1.9) 22.6 TPSD	0	1,883	0	0	0	N.D.
75* (4.2-23)	Excess Mine Water to Aeration Pond	8,330	0	0	0	4,170	N.D.	4,165,000
76* (4.2-23)	Aerated Water to Discharge	8,330	0	0	0	4,170	N.D.	4,165,000
77	Feedwater to Waste Heat Boiler	2,120	0	0	0	50	N.D.	1,060,000
78*	Total Processed Shale Moistening Water	5,624	0	0	0	N.D.	N.D.	2,812,000
79	Cooling Water to Lurgi Oil Recovery	325	0	0	0	163	0	162,500
80	Cooling Water to Naphtha Recovery	5	0	0	0	2.5	0	2,500
81	Steam Condensate from Naphtha Stripper	20	0	0	0	0	0	10,000
82	Cooling Water to Compression Cooling	8	0	0	0	4	0	4,000
83	Cooling Water to DEA-TEG Treatment	66	0	0	0	33	0	33,000
84	Steam Condensate from DEA-TEG Treatment	260	0	0	0	0	0	130,200
85	Steam Condensate from Stretford	2	0	0	0	0	0	1,000
86	Cooling Water Makeup to Stretford	2	0	0	0	1	0	1,000
87	Process Water Makeup to Stretford	3	0	0	0	0	0	1,500
88*	Humidified Air Cooler Blowdown	661	0	0	0	496	0	330,500
89	Cooling Water to Ammonia Recovery	1,080	0	0	0	540	0	540,000
90*	Water for Dust Palliatives	1,568	0	0	0	784	0	784,000
91*	Processed Shale Revegetation Water	649	0	0	0	325	0	324,500

(Continued)

TABLE 4.1-4 (cont.)

Stream Number (Table No.)	Stream Description	Mass Flow, gpm (10^3 lb/hr)	Components, lb/hr					
			CO_2	NH_3	H_2S	TDS & TSS	Organics	H_2O
92* (4.2-3, 4.2-4)	Raw Shale Leachate	N.D.	0	4.6 µg/l	0	3,490 mg/l	13 mg/l	N.D.
93*	Storm Runoff	150	0	0	0	N.D.	N.D.	75,000
94	Boiler Feedwater Makeup	43	0	0	0	1	0	21,500
95*	Service and Fire Water	43	0	0	0	22.5	0	21,500
96*	Mine Water Clarifier Sludge	165	0	0	0	N.D.	N.D.	82,500
97	Water to Cooling Tower Makeup Treatment	2,676	0	0	0	1,340	0	1,338,000
98	Treated Water to Cooling Tower	2,676	0	0	0	1,340	0	1,338,000
99*	Potable/Sanitary Water	26	0	0	0	13	0	13,000
101*	Used Sanitary Water to Municipal Treatment	18	0	0	0	N.D.	N.D.	9,000
102*	Treated Sanitary Water	18	0	0	0	N.D.	0	9,000
103*	Sanitary Water Treatment Sludge	N.D.	0	0	0	N.D.	N.D.	N.D.
104*	Boiler Feedwater Treatment Concentrate	11	0	0	0	N.D.	N.D.	5,500
105*	Cooling Tower Blowdown	1,123	0	0	0	842	N.D.	561,500
106*	Cooling Tower Drift	9	0	0	0	5	0	4,500
108*	Equalization Pond Discharge to Processed Shale Moistening	2,525	0	0	0	N.D.	N.D.	1,262,500
109*	Clarified Mine Water to Processed Shale Moistening	2,912	0	0	0	1,096	0	1,456,000
111*	Aerated Pond Sludge	N.D.	0	0	0	N.D.	N.D.	N.D.

* Indicates streams that come into contact with the environment.

**N.D. = Not determined; 0 = Estimated to be insignificant (less than 1 lb).

Source: DRI estimates based on information from Gulf Oil Corp. and Standard Oil Co. (Indiana), March 1976, and Rio Blanco Oil Shale Co., February 1981.

TABLE 4.1-5. INVENTORY OF SOLID STREAMS

Stream Number (Table No.)	Description of Stream	Mass Flow, 10^3 lb/hr	Components of Concern	Mass Flow of Component, lb/hr	Remarks[a]
1* (4.2-2)	Raw Shale Feed	9,799	Particulates	118,100	Dust collection and suppression are employed to minimize the particulate emissions from the raw shale handling operations.
2*	Subore	992	Particulates	520	The subore is crushed and disposed of with the processed shale. The dust from crushing is controlled with baghouses.
3*	Overburden	5,175	Particulates	2,730	The overburden is crushed and disposed of with the processed shale. Dust from crushing is controlled with baghouses.
22* (4.2-2)	Baghouse Dusts	118.1	--[b]	--	This dust is collected from raw shale handling operations and combined with the raw shale for retorting.
29* (4.2-5, 4.2-6, 4.2-7)	Processed Shale	9,733	Particulates Leachable Salts	2,820 280,000	The processed shale is properly moisturized to reduce dust emissions. Proper compaction should reduce water permeability, hence leaching of salts.
43	Oily Dust	78.6	Adsorbed Oil Residual Organics	N.D.[c] ~1,800	This dust is obtained from heavy oils dedusting. It is incinerated in the lift pipes along with the bulk of the processed shale.
68	Caustic (NaOH)	0.3	--	--	Caustic is added to the ammonia recovery process to make up the reagent losses as well as to release the fixed ammonia.

* Indicates streams that come into contact with the environment.

[a] The remarks indicate the stream disposition. The controls and treatments applied to the streams are those proposed for the Lurgi-Open Pit technologies.

[b] Dashes (--) indicate no known components of concern.

[c] N.D. = Not determined.

Source: DRI estimates based on information from Gulf Oil Corp. and Standard Oil Co. (Indiana), March 1976, and Rio Blanco Oil Shale Co., February 1981.

TABLE 4.1-6. COMPOSITIONS OF SOLID STREAMS

Stream Number (Table No.)	Stream Description	Mass Flow, 10^3 lb/hr	Components,[a] 10^3 lb/hr					
			H	O	N	C	S	H_2O
1* (4.2-2)	Raw Shale Feed	9,799	131	73	40	1,025	98	261
2*	Subore	992	N.D.[b]	N.D.	N.D.	N.D.	N.D.	N.D.
3*	Overburden	5,175	N.D.	N.D.	N.D.	N.D.	N.D.	N.D.
22* (4.2-2)	Baghouse Dusts	118.1	2	1	0.2	12	1	3
29* (4.2-5, 4.2-6, 4.2-7)	Processed Shale	9,733	0[b]	2	8	24	91	1,820
43	Oily Dust	78.6	0	0.02	0.06	0.19	0.73	0
68	Caustic (NaOH)	0.3	0	0	0	0	0	0

* Indicates streams that come into contact with the environment.

[a] Elements reported for organic portion of materials, except for sulfur which is total.

[b] N.D. = Not determined; 0 = Estimated to be insignificant (less than 1 lb).

Source: DRI estimates based on information from Gulf Oil Corp. and Standard Oil Co. (Indiana), March 1976, and Rio Blanco Oil Shale Co., February 1981.

4.2 MAJOR STREAM COMPOSITIONS

Much of the significant data for the Lurgi-Ruhrgas retorting process have been proprietary in the past and largely remain so at present. The limited information that is available has been extracted from Rio Blanco Oil Shale Company's (RBOSC) Modification to the Detailed Development Plan (DDP) (February 1981) and communications with RBOSC and Lurgi Kohle und Mineralötechnik GmbH personnel. Some generalized information on the retorting technology is also published and this was used when appropriate (Marnell, September 1976; Schmalfeld, July 1975).

In the following sections, major streams generated from different plant operations (see Section 3) are listed along with their detailed compositions. Material balances for selected streams (both before and after treatment) are also presented. When detailed information on stream compositions or performance of a control technology was not available, calculations were made on the basis of engineering analysis.

4.2.1 Material Balance

The material balance for retorting 23 gpt oil shale by the Lurgi-Ruhrgas process is presented in Table 4.2-1. This balance is for the retorting process only. The amount of raw shale retorted (119,000 TPSD) is derived from the original DDP (Gulf Oil Corp. and Standard Oil Co. [Indiana], March 1976). The combustion air, processed shale, quenching and moisturizing water, net retort gas, and flue gas quantities have been calculated using the modified DDP (Rio Blanco Oil Shale Co., February 1981) for the Lurgi demonstration project. Amounts of oil, naphtha, and retort gas have been estimated assuming a 100% Fischer assay oil yield and, also, by material and elemental balances. After pyrolizing the shale, the amount of coke remaining is insufficient for raising the recycle shale stream to the desired temperature of about 1,240°F; therefore, a portion of the retort gas (before naphtha removal) is added to the lift pipes as supplemental fuel.

4.2.2 Raw Oil Shale

The exact composition of the raw shale was not available. Therefore, an estimation was made using the published analyses of different grades of Green River oil shale and its kerogen (Stanfield, et al., 1951). The estimates are fairly representative of expected values and are further strengthened by good overall material and elemental balances. Derived composition for the raw shale is presented in Table 4.2-2.

Raw Shale Leachate--

Recently, some literature on leachates from Colorado oil shales has been published (McWhorter, 1980; Rio Blanco Oil Shale Co., March 1981). The results of laboratory column leaching experiments from the first reference are presented in Table 4.2-3. The second reference provided field lysimeter study results from Tract C-a run-of-mine stockpile tests, and these are shown in Table 4.2-4.

TABLE 4.2-1. GROSS MATERIAL BALANCE FOR RETORT AND SHALE BURNER

Material In		Flow, 10^3 lb/hr
Raw Shale		9,917
Air		5,439
Makeup Water[a]		2,812
	Total In	18,168
Material Out		**Flow, 10^3 lb/hr**
Processed Shale[b]		9,733
Retort Gas (net, naphtha-free)[c]		122
Gas Liquor		298
Product Oil		793
Naphtha (net)[d]		27
Flue Gas		7,195
	Total Out	18,168

[a] The makeup water includes 992 × 10^3 lb/hr for processed shale quenching and 1,820 × 10^3 lb/hr for processed shale moisturizing to a moisture content of approximately 19% by weight.

[b] The processed shale is burned (after the lift pipes) and includes the moisturizing water. The processed shale quantity on a dry basis would be 7,913 × 10^3 lb/hr.

[c] The net retort gas quantity is that remaining after subtracting 87.4 × 10^3 lb/hr of the gas used in the lift pipes.

[d] The net naphtha quantity is that remaining after subtracting 19.6 × 10^3 lb/hr of the naphtha used along with the retort gas in the lift pipes.

Source: DRI estimates based on data from Rio Blanco Oil Shale Co., February 1981.

TABLE 4.2-2. COMPOSITION OF RAW SHALE*
(Streams 1, 22)

Component	Weight Percent	Mass Flow, 10^3 lb/hr	Flow, 10^3 lb-moles/hr
Raw Shale	100.00	9,917	--
Hydrogen (organic)	1.34	133	133.0
Moisture	2.66	264	14.7
Oxygen (organic)	0.75	74	4.6
Nitrogen (organic)	0.40	40	2.9
Carbon (organic)	10.46	1,037	86.4
Sulfur (total)	1.00	99	3.1

* Based on 65,167 BPSD crude shale oil at 100% Fischer assay yield, with 23 gpt oil shale. Baghouse dusts are included.

Source: DRI estimates based on information from Stanfield, et al., 1951.

TABLE 4.2-3. LABORATORY COLUMN LEACHATES FROM SOME COLORADO RAW OIL SHALES
(Stream 92)

Component	Unit	USBM Raw Shale (Saline Zone)	Colony Raw Shale	C-a R-5/Mahog. Shale	C-a Composite	Raw Colony Shale (Weathered)	Colony Soil	C-b Soil	Drinking Water Criteria
Al	mg/l	0.34 - 7.54	0.05 - 0.75	0.3 - 3.53	<0.05 - 0.69	<0.05 - 0.17	<0.05 - 0.37	<0.05	--
As	"	<0.005	<0.005	<0.005	<0.005	<0.005	<0.005	<0.005	0.05
B	"	0.24 - 43	<0.025 - 2.75	<0.025 - 0.59	<0.025 - 1.97	0.12 - 0.365	0.47 - 0.76	0.65 - 0.985	--
Ba	"	0.061 - 0.17	0.07 - 0.48	0.088 - 0.27	0.027 - 0.22	0.088 - 0.495	0.12 - 0.57	0.038 - 0.240	1.0
Be	"	<0.025	<0.025	<0.025	<0.025	<0.025	<0.025	<0.025	--
Ca	"	36 - 750	40 - 1,550	180 - 1,510	18 - 970	25 - 500	40 - 960	6.5 - 330	--
Cl	"	<1.0 - 560	1.1 - 22	1.9 - 300	0.3 - 130	0.8 - 71	0.1 - 200	1.8 - 520	--
CO_3	"	0.1 - 1.1	0.03 - 1.6	<0.1 - 346	0.3 - 0.7	0.12 - 1.03	0.17 - 2.9	1.0 - 8.8	--
Cr	"	<0.025 - 0.68	<0.025 - 0.04	0.022 - 0.034	<0.025 - 0.043	<0.025	<0.025	<0.025 - 0.069	0.05
Cu	"	<0.025 - 0.30	<0.025 - 0.41	<0.025 - 0.69	<0.025 - 0.44	<0.039 - 0.31	0.075 - 0.38	<0.025 - 0.28	1.0
EC	μmhos/cm	280 - 13,000	240 - 5,400	1,900 - 37,000	125 - 8,200	240 - 4,200	370 - 9,000	840 - 3,000	--
F	mg/l	9.5 - 75	4.0 - 7.2	0.8 - 65	<0.5 - 3.0	4.0 - 8.7	1.2 - 10	4.5 - 25	1.8
Fe	"	0.01 - 1.8	<0.03 - 0.89	<0.1	<0.1	<0.03 - 0.08	<0.01 - 0.52	0.1 - 0.42	0.3
HCO_3	"	83.1 - 321	50 - 558	3.0 - 403	82 - 1,026	136 - 233	152 - 480	481 - 846	--
Hg	"	<0.0001 - 0.0035	<0.0005	<0.0001	<0.0001	<0.0005	<0.0005	<0.0001	0.002
K	"	1.1 - 22	1.7 - 59	8.2 - 640	0.4 - 34	0.83 - 57	25 - 270	1.3 - 22	--
Li	"	0.02 - 3.1	0.02 - 0.151	0.02 - 0.11	<0.02 - 0.79	0.004 - 0.02	0.03 - 0.47	0.02 - 0.08	--
Mg	"	6.7 - 1,050	5.5 - 140	0.675 - 108	4.9 - 820	17 - 365	17 - 1,450	2.6 - 145	--
Mn	"	0.075 - 3.2	0.074 - 2.74	<0.05 - 0.35	<0.05 - 0.40	0.05 - 0.11	<0.05 - 0.97	<0.05 - 0.16	0.05
MO	"	0.09 - 0.87	0.09 - 0.65	0.10 - 5.18	0.10 - 2.2	0.075 - 0.74	<0.05 - 0.84	<0.05 - 0.43	--
Na	"	<25 - 1,430	5.8 - 145	27 - 7,710	4.3 - 1,240	14 - 350	3.8 - 340	210 - 2,050	--
Ni	"	<0.025 - 0.60	<0.05 - 0.10	0.047 - 0.085	<0.05 - 0.16	<0.05 - 0.06	<0.05 - 0.07	<0.05 - 0.075	--
NO_3	"	<1.25 - 40	0.9 - 25	4 - 172	<0.5 - 140	<0.3 - 245	<0.2 - 180	1.4 - 30	10
Pb	"	<0.04 - 1.9	<0.05 - 0.64	<0.05 - 0.83	<0.05 - 0.77	0.12 - 0.35	0.12 - 0.38	0.07 - 0.31	0.05
pH	--	6.8 - 8.06	7.06 - 8.18	6.93 - 11.98	7.03 - 7.99	6.93 - 8.11	7.1 - 8.2	7.43 - 8.49	--
Se	mg/l	<0.01	<0.01	<0.005	<0.005	<0.01	<0.01	<0.005	0.01
Si	"	1.65 - 9.7	2.12 - 10.58	1.2 - 23.28	5.8 - 19.58	6.71 - 14.72	8.0 - 16.8	11.0 - 20.7	--
Sn	"	<0.025 - 1.28	0.12 - 0.67	--	--	0.041 - 0.67	<0.025 - 1.37	--	--
SO_4	"	20 - 5,700	28 - 5,150	5 - 6,600	7.9 - 6,100	15 - 2,650	60 - 4,200	23 - 860	250
TDS	"	70 - 13,300	110 - 7,160	610 - 30,130	164 - 9,450	120 - 4,760	250 - 7,450	1,050 - 3,760	500
Zn	"	0.01 - 6.8	<0.02 - 0.68	<0.01 - 0.09	<0.02 - 1.5	0.07 - 0.3	0.01 - 0.65	0.04 - 0.35	5.0

Source: McWhorter, 1980.

TABLE 4.2-4. LEACHATE WATER QUALITY DATA FROM THE TRACT C-a RUN-OF-MINE STOCKPILE

(Stream 92)

Constituent	Concentration*	Constituent	Concentration*
Alkalinity (mg/l as $CaCO_3$)	48.0	Nitrate-Nitrite (as N)	45.0
Aluminum	10.0	Ammonia (as N)	4.6
Arsenic	2.0	Kjeld-N (as N)	9.1
Barium	100.0	DON (as N)	4.5
Beryllium	0.0	Total Nitrogen (as N)	54.0
Boron	150.0	pH (field)	7.7
Cadmium	0.0	pH (lab)	7.1
Calcium (mg/l)	440.0	Phenols	3.0
DOC (mg/l)	13.0	Total Phosphorus (mg/l)	0.0
Chloride (mg/l)	29.0	Potassium (mg/l)	3.5
Chromium	0.0	Potassium 40 (pc/l)	2.6
Cobalt	100.0	TDS (calculated) (mg/l)	3,490.0
Copper	8.0	SAR	1.6
Fluoride (mg/l)	2.4	Selenium	0.0
Hardness (noncarbonate)		Silica (mg/l)	4.6
(mg/l)	2,300.0	Sodium (mg/l)	180.0
Hardness (total) (mg/l)	2,300.0	Sodium (%)	14.0
Iron	60.0	Spec. cond. (field)	
Lead	0.0	(µmhos/cm)	3,950.0
Lithium	50.0	Spec. cond. (lab)	
Magnesium (mg/l)	300.0	(µmhos/cm)	3,877.0
Manganese	220.0	Strontium	3,000.0
Mercury	0.0	Sulfate (mg/l)	2,300.0
Molybdenum	270.0	Vanadium	4.0
Nickel	37.0	Zinc	50.0

* All concentrations are expressed in µg/l, unless listed otherwise, and apply to the dissolved fraction only.

Source: Rio Blanco Oil Shale Co., March 1981.

4.2.3 Processed Shale

The quantity and composition of the processed shale, derived by material and elemental balances, are presented in Table 4.2-5. Due to burning of the processed shale in the lift pipes and extensive recycling to the retort, the residual organic matter is fairly low. The moisturizing water amounts to approximately 23% of the dry processed shale weight. Major inorganic elements in the processed shale obtained from a retorting test on oil shale from Tract C-a are presented as their oxides in Table 4.2-6. Some physical properties of the processed shale have also been determined and are presented in Table 4.2-7. Due to partial calcination in the lift pipes, the processed shale has good cementitious properties. The unconfined compactive strength, at optimum moisture content and curing period, is high and permeability is low.

TABLE 4.2-5. COMPOSITION OF THE PROCESSED MOISTURIZED SHALE
(Stream 29)

Component	Weight Percent	Mass Flow, 10^3 lb/hr	Flow, 10^3 lb-moles/hr
Retorted Shale (moisturized)	100.00	9,733	--
Moisture	18.70	1,820	101.1
Oxygen (organic)	0.02	2	0.1
Nitrogen (organic)	0.08	8	0.6
Carbon (organic)	0.25	24	2.0
Sulfur (total)	0.93	91	2.8

Source: DRI estimates based on information from Rio Blanco Oil Shale Co., February 1981.

Processed Shale Leachate--

The results from column leaching of processed shale are given in Table 4.2-8 (Woodward-Clyde Consultants, October 13, 1980). Some soluble elements are reported as their oxides. As seen in Table 4.2-7, properly moistened and compacted processed shale has low permeability; therefore, actual field leaching may not be represented by laboratory column leaching experiments.

TABLE 4.2-6. INORGANIC ANALYSIS OF THE PROCESSED SHALE
(Stream 29)

Component	Weight Percent
Silicon Dioxide	46.00
Iron Oxides	4.40
Aluminum Oxide	12.70
Calcium Oxide	22.40
Magnesium Oxide	4.80
Sulfate	3.80
Sodium Oxide	3.20
Potassium Oxide	2.70
Carbonate	4.40
Chloride	0.08
Loss on Ignition	4.60

Source: Woodward-Clyde Consultants, October 13, 1980. The results are for processed shale from a retorting test on Tract C-a oil shale.

TABLE 4.2-7. PHYSICAL PROPERTIES OF PROCESSED SHALE
(Stream 29)

		Gradation						Compaction		Permeability ft/yr			Shear Strength					
													Triaxial Shear				Unconfined	
Test Condition	Maximum Particle Size, in.	% Clay (0.005 mm)	% Silt (-200 mesh - 0.005 mm)	% Sand (4 - 200 mesh)	% Gravel (+4 mesh)	Remarks	Specific Gravity	% Optimum Moisture	Maximum Density, PCf	Load, 50 psi	Load, 100 psi	Load, 200 psi	Friction Coefficient, Tan Ø	Internal Angle of Friction, °	Cohesion, psi	Saturated Cohesion, psi	Days Cured @ 120°F	Unconfined Compression, psi
As Received	3/8	17.1	47.8	33.4	1.7		2.83											
	3/8	15.3	46.4	36.6	1.7		2.84											
Compacted	--	16.2	48.2	35.6	0.0	Initial		30.3	85.6	0.002	0.003	--	0.69	34.5	22.2	7.6	0	28.8
1/2 D 698	--	17.6	44.2	38.2	0.0	After Compaction											0	33.1
(6,200 ft-lbs)	--	15.5	37.9	46.6	0.0	After Triaxial Shear Test											7	590.7
		15.5	37.9	46.6	0.0												7	785.9
																	14	575.0
																	14	682.4
																	28	--
																	28	--
Compacted	--	16.2	48.2	35.6	0.0	Initial		28.5	88.2	0.003	0.005	--	0.62	32.0	33.3	13.9	0	38.5
D 698	--	18.8	44.2	37.0	0.0	After Compaction											0	42.5
(12,375 ft-lbs)	--	8.8	42.5	48.7	0.0	After Triaxial Shear Test											7	874.5
																	7	865.5
																	14	940.8
																	14	912.1
																	28	1,222.2
																	28	1,222.1
Compacted	--	16.2	48.2	35.6	0.0	Initial		23.2	96.8	0.001	0.001	--	0.80	38.5	41.0	27.8	0	378.5
D 1557	--	0.1	56.3	43.6	0.0	After Compaction											0	350.8
(56,250 ft-lbs)	--	12.8	45.9	41.3	0.0	After Triaxial Shear Test											7	971.1
																	7	1,182.6
																	14	986.5
																	14	1,081.0
																	28	--
																	28	--

Source: Woodward-Clyde Consultants, October 13, 1980. The results are for processed shale from a retorting test on Tract C-a oil shale.

TABLE 4.2-8. ANALYSIS OF LEACHATE FROM THE PROCESSED SHALE

Component	Concentration, mg/l
Silicon Dioxide	18
Iron Oxides	<0.01
Aluminum Oxide	<0.1
Calcium Oxide	1,080
Magnesium Oxide	102
Sodium Oxide	337
Potassium Carbonate	37
Carbonate	90
Bicarbonate	<0.1
Chloride	28
Sulfate	1,810
Hydroxide	222
Total Dissolved Solids	3,530
pH = 11.4	

Source: Woodward-Clyde Consultants, October 13, 1980. The results are for processed shale from a retorting test on Tract C-a oil shale.

4.2.4 Crude Shale Oil

The composition of vapors from the Lurgi retorts is indicated in Table 4.2-9. The Lurgi retorts also include three condensation-absorption towers; consequently, a product breakdown of the condensable hydrocarbons occurs, forming heavy, middle and light oil fractions. The properties of the individual oil fractions are indicated in Table 4.2-10. Since the naphtha fraction is still contained in the gas phase, it is not included in the table. The physical properties for each oil fraction have been estimated using the oil distillation data published in RBOSC's modified DDP (Rio Blanco Oil Shale Co., February 1981). The composition for the combined shale oil has been calculated by material and elemental balances and from data provided by Occidental Oil Shale, Inc.

4.2.5 Retort Gas

The heavy oils and most of the entrained dust in the retort vapors are eliminated in the first condensation tower. The middle oil fractions are condensed in the second tower by reducing the vapor temperature to 150°F by wet cooling. The light oils, naphtha, water, and noncondensable gases remain

TABLE 4.2-9. COMPOSITION OF RETORT VAPORS
(Stream 26)

Component	MWt	Mass %	Mole %	Mass Flow, 10^3 lb/hr	Flow, lb-moles/hr
H_2	2	0.37	8.20	4.97	2,486.6
CO	28	0.53	0.84	7.14	255.0
CO_2	44	7.31	7.39	98.46	2,237.7
N_2	28	0.53	0.84	7.14	255.0
NH_3	17	0.14	0.37	1.92	113.2
H_2S	34	0.09	0.12	1.20	35.2
SO_2	64	0.02	0.01	0.26	4.0
CH_4	16	1.41	3.91	18.95	1,184.1
C_2H_4	28	0.91	1.44	12.24	437.2
C_2H_6	30	0.93	1.38	12.57	419.0
C_3H_6	42	1.16	1.23	15.68	373.4
C_3H_8	44	0.63	0.63	8.42	191.3
C_4H_8	56	1.02	0.81	13.77	245.9
C_4H_{10}	58	0.31	0.24	4.23	72.9
C_4^+	79.4	1.77	0.99	23.87	300.6
Light Oils	114	15.28	5.96	205.77	1,805.0
Middle Oils	166	30.69	8.22	413.24	2,489.4
Heavy Oils	274	14.59	2.37	196.51	717.2
Miscellaneous HC	132.6	0.02	0.01	0.24	1.8
H_2O	18	22.28	55.03	300.05	16,669.4
TOTAL		99.99	99.99	1,346.63*	30,293.9
MWt		44.45			

* In addition, approximately 78,600 lb/hr of processed shale dust are entrained in the retort vapors. The presence of COS and other organic sulfur compounds has not been determined.

Source: DRI estimates based on data from Rio Blanco Oil Shale Co., February 1981, and provided by Occidental Oil Shale, Inc.

TABLE 4.2-10. PROPERTIES OF NAPHTHA-FREE SHALE OIL
(Streams 36, 38, 42)

Component	Boiling Point, °F	Volume Percent (BPSD)	Gravity, °API	Weight Percent	Mass Flow, 10^3 lb/hr	Flow, lb-moles/hr
Heavy Oils (Stream 42)	~360-920	23.65 (14,260)	18.5	24.78	196.51	717.20
Middle Oils (Stream 38)	~270-920	53.27 (32,120)	30.0	51.87	411.33	2,477.90
Light Oils* (Stream 36)	~290-510	23.08 (13,920)	24.0	23.35	185.16	1,624.20
TOTAL SHALE OIL	~270-920	100.00 (60,300)	25.5	100.00	793.00	4,819.30
Composition						
Hydrogen	--	--	--	11.47	91.00	91.00
Oxygen	--	--	--	1.37	11.00	0.69
Nitrogen	--	--	--	2.05	16.00	1.14
Carbon	--	--	--	84.2	668.00	55.67
Sulfur	--	--	--	0.86	7.00	0.22

* The light oils are stabilized. API gravity for the light oils before stabilization is 36.5°.

Source: DRI estimates based on data from Rio Blanco Oil Shale Co., February 1981, and provided by Occidental Oil Shale, Inc.

in the vapor phase (stream 33). The composition of this stream, as it enters the third tower, is estimated in Table 4.2-11. The light oils and gas liquor are condensed in the third tower by further reducing the temperature of the gases to about 90°F. Most of the ammonia and polar organic compounds are scrubbed out in the gas liquor, but naphtha still remains in the retort gas (stream 34). A portion of this gas is used as a fuel in the lift pipes, while the remainder is sent to the lean oil absorber for the recovery of naphtha. The net retort gas composition, after condensing the light oils and gas liquor, is presented in Table 4.2-12. In the lean oil absorber, the retort gas is scrubbed with the light oils obtained earlier. Heavier hydrocarbons (naphtha) in the gas are absorbed in the lean oil, while the noncondensable gases (stream 45) pass on for further purification. The composition of noncondensable, or naphtha-free, retort gas is given in Table 4.2-13. The rich light oils containing the absorbed hydrocarbons are then steam stripped to recover the naphtha (stream 46). The estimated composition of the naphtha is presented in Table 4.2-14. The lean light oils are recycled back to the lean oil absorber.

Since the retort gas is being prepared for sale, it undergoes further purification. First, most of the moisture and residual ammonia and sulfur dioxide are removed by compressing and cooling the naphtha-free retort gas. The composition of this gas (stream 48) is presented in Table 4.2-15. The condensed water is combined with the gas liquor for subsequent treatment. The compressed gas is then treated for acid gas removal in a diethanolamine (DEA) absorber. A majority of the carbon dioxide and hydrogen sulfide is absorbed in the amine solution, and the retort gas exits the absorber in fairly pure form. Table 4.2-16 presents the estimated composition of the amine-treated retort gas. Due to the removal of most of the CO_2, the heating value of the gas is almost doubled. As a final step, the residual moisture in the gas is removed by absorption in triethylene glycol, and the dried fuel gas (stream 57), as represented in Table 4.2-17, is sent to the pipeline.

The rich amine solution containing absorbed CO_2 and H_2S is regenerated by steam stripping, and the desorbed acid gases (stream 58) are treated for sulfur recovery in the Stretford unit. Table 4.2-18 presents the material balance for the Stretford unit. The Stretford-treated acid gases (stream 63) are virtually free of sulfur compounds; thus, they are emitted to the atmosphere. Since the oxidizer vent gas (stream 64) is primarily air, it is used as a source of combustion air to the lift pipes, rather than being released directly to the environment.

4.2.6 Flue Gas

Flue gas (stream 31) from the Lurgi process results from the combustion of organic residue on the processed shale and the fuel gas in the lift pipes. The quantity and composition of the flue gas, calculated from material and elemental balances, are presented in Table 4.2-19. The calculated composition is consistent with the data reported in the modified DDP (Rio Blanco Oil Shale Co., February 1981). Various plant streams, such as the ammonia recovery overhead vapors and Stretford oxidizer vent gas, are also added to the lift pipes. Since these streams do not have a significant impact on the

TABLE 4.2-11. COMPOSITION OF RETORT GAS AFTER MIDDLE OILS SCRUBBER
(Stream 33)

Component	MWt	Mass %	Mole %	Mass Flow, 10^3 lb/hr	Flow, lb-moles/hr
H_2	2	0.67	9.17	4.97	2,486.6
CO	28	0.97	0.94	7.14	255.0
CO_2	44	13.33	8.26	98.46	2,237.7
N_2	28	0.97	0.94	7.14	255.0
NH_3	17	0.26	0.42	1.92	113.2
H_2S	34	0.16	0.13	1.20	35.2
SO_2	64	0.03	0.01	0.26	4.0
CH_4	16	2.57	4.37	18.95	1,184.1
C_2H_4	28	1.66	1.61	12.24	437.2
C_2H_6	30	1.70	1.55	12.57	419.0
C_3H_6	42	2.12	1.38	15.68	373.4
C_3H_8	44	1.14	0.71	8.42	191.3
C_4H_8	56	1.86	0.91	13.77	245.9
C_4H_{10}	58	0.57	0.27	4.23	72.9
C_4^+	79.4	3.23	1.11	23.87	300.6
Light Oils	114	27.85	6.66	205.77	1,805.0
Middle Oils	166	0.26	0.04	1.91	11.5
Miscellaneous HC	132.6	0.03	0.01	0.24	1.8
H_2O	18	40.61	61.52	300.05	16,669.4
TOTAL		99.99	100.01	738.79	27,098.8
MWt		27.26			

Source: DRI estimates based on data from Rio Blanco Oil Shale Co., February 1981.

TABLE 4.2-12. COMPOSITION OF RETORT GAS AFTER LIGHT OILS SCRUBBER
(Stream 34)

				Mass Flow, 10^3 lb/hr		Net
Component	MWt	Mass %	Mole %	Total	Net*	lb-moles/hr
H_2	2	1.94	27.29	4.97	2.89	1,447.3
CO	28	2.79	2.80	7.14	4.16	148.4
CO_2	44	37.57	23.99	96.18	55.98	1,272.3
N_2	28	2.79	2.80	7.14	4.16	148.4
NH_3	17	0.02	0.03	0.05	0.03	1.7
H_2S	34	0.47	0.39	1.20	0.70	20.5
SO_2	64	0.01	0.01	0.02	0.02	0.3
CH_4	16	7.40	12.99	18.95	11.03	689.2
C_2H_4	28	4.78	4.80	12.24	7.13	254.5
C_2H_6	30	4.91	4.60	12.57	7.32	243.9
C_3H_6	42	6.13	4.10	15.68	9.13	217.3
C_3H_8	44	3.29	2.10	8.42	4.90	111.3
C_4H_8	56	5.38	2.70	13.77	8.01	143.1
C_4H_{10}	58	1.65	0.80	4.23	2.46	42.4
C_4^+	79.40	9.32	3.30	23.87	13.89	174.9
Heavy Ends	81.69	8.79	3.02	22.51	13.10	160.4
H_2O	18	2.75	4.30	7.05	4.10	228.0
TOTAL		99.99	100.02	255.99	149.01	5,303.9
MWt		28.09				
Total H (excluding H_2O)		10.24		26.21	15.26	15,260
Total O (excluding H_2O)		28.92		74.04	43.10	2,694
Total N		2.80		7.18	4.18	299
Total C		54.82		140.35	81.69	6,808
Total S		0.45		1.14	0.67	21

* The net amount is treated for naphtha recovery. The remainder is sent to the lift pipes as supplemental fuel.

Source: DRI estimates based on data from Rio Blanco Oil Shale Co., February 1981.

TABLE 4.2-13. COMPOSITION OF NAPHTHA-FREE RETORT GAS
(Stream 45)

Component	MWt	Mass %	Mole %	Mass Flow, 10^3 lb/hr	Flow, lb-moles/hr
H_2	2	2.38	29.22	2.89	1,447.3
CO	28	3.41	3.00	4.16	148.4
CO_2	44	45.99	25.68	55.98	1,272.3
N_2	28	3.41	3.00	4.16	148.4
NH_3	17	0.02	0.03	0.03	1.7
H_2S	34	0.57	0.41	0.70	20.5
SO_2	64	0.02	0.01	0.02	0.3
CH_4	16	9.06	13.91	11.03	689.2
C_2H_4	28	5.85	5.14	7.13	254.5
C_2H_6	30	6.01	4.92	7.32	243.9
C_3H_6	42	7.50	4.39	9.13	217.3
C_3H_8	44	4.02	2.25	4.90	111.3
C_4H_8	56	6.58	2.89	8.01	143.1
C_4H_{10}	58	2.02	0.86	2.46	42.4
H_2O	18	3.15	4.30	3.83	212.9
TOTAL		99.99	100.01	121.73	4,953.5
MWt		24.57			
Total H (excluding H_2O)		9.81		11.94	11,942
Total O (excluding H_2O)		35.40		43.10	2,694
Total N		3.43		4.18	299
Total C		47.66		58.03	4,835
Total S		0.55		0.67	21
Heating Value, LHV Btu/lb (Btu/SCF)			9,794 (616)		

Source: DRI estimates based on data from Rio Blanco Oil Shale Co., February 1981.

TABLE 4.2-14. COMPOSITION OF NAPHTHA
(Stream 46)

Component	MWt	Mass %	Mole %	Mass Flow, 10^3 lb/hr	Flow, lb-moles/hr
C_5H_{12}	72	25.22	27.25	6.88	95.5
C_6H_6	78	1.52	1.51	0.41	5.3
C_6H_{14}	86	18.39	16.64	5.01	58.3
C_7H_{16}	100	5.83	4.54	1.59	15.9
Light Oils	78	43.97	43.86	11.99	153.7
Middle Oils	166	4.08	1.91	1.11	6.7
H_2O	18	0.99	4.28	0.27	15.0
TOTAL		100.00	99.99	27.26	350.4
MWt		77.81			
Total H		12.17		3.32	3,318
Total C		86.84		23.68	1,973

Source: DRI estimates based on data from Rio Blanco Oil Shale Co., February 1981.

TABLE 4.2-15. COMPOSITION OF RETORT GAS AFTER COMPRESSION
(Stream 48)

Component	MWt	Mass %	Mole %	Mass Flow, 10^3 lb/hr	Flow, lb-moles/hr
H_2	2	2.45	30.51	2.89	1,447.3
CO	28	3.52	3.13	4.16	148.4
CO_2	44	47.45	26.81	55.96	1,271.9
N_2	28	3.52	3.13	4.16	148.4
NH_3	17	0.0014	0.0021	0.002	0.1
H_2S	34	0.59	0.43	0.70	20.5
CH_4	16	9.35	14.53	11.03	689.2
C_2H_4	28	6.04	5.37	7.13	254.5
C_2H_6	30	6.20	5.14	7.32	243.9
C_3H_6	42	7.74	4.58	9.13	217.3
C_3H_8	44	4.15	2.35	4.90	111.3
C_4H_8	56	6.80	3.02	8.01	143.1
C_4H_{10}	58	2.09	0.89	2.46	42.4
H_2O	18	0.08	0.11	0.10	5.4
TOTAL		100.00	100.00	117.95	4,743.7
MWt		24.86			
Total H (excluding H_2O)		10.12		11.94	11,937
Total O (excluding H_2O)		36.52		43.08	2,692
Total N		3.52		4.16	297
Total C		49.18		58.01	4,834
Total S		0.56		0.66	21
Heating Value, LHV Btu/lb (Btu/SCF)			10,010 (656)		

Source: DRI estimates based on data from Rio Blanco Oil Shale Co., February 1981.

TABLE 4.2-16. COMPOSITION OF RETORT GAS AFTER AMINE ABSORBER
(Stream 56)

Component	MWt	Mass %	Mole %	Mass Flow, 10^3 lb/hr	Flow, lb-moles/hr
H_2	2	4.68	41.77	2.89	1,447.3
CO	28	6.71	4.28	4.16	148.4
CO_2	44	1.00	0.41	0.62	14.1
N_2	28	6.71	4.28	4.16	148.4
NH_3	17	0.0027	0.0029	0.0017	0.1
H_2S	34	0.0005	0.0003	0.0003	0.01
CH_4	16	17.82	19.89	11.03	689.2
C_2H_4	28	11.52	7.35	7.13	254.5
C_2H_6	30	11.82	7.04	7.32	243.9
C_3H_6	42	14.75	6.27	9.13	217.3
C_3H_8	44	7.91	3.21	4.90	111.3
C_4H_8	56	12.95	4.13	8.01	143.1
C_4H_{10}	58	3.97	1.22	2.46	42.4
H_2O	18	0.14	0.14	0.09	4.8
TOTAL		99.98	99.99	61.90	3,464.8
MWt		17.86			
Total H (excluding H_2O)		19.22		11.90	11,896
Total O (excluding H_2O)		4.58		2.83	177
Total N		6.72		4.16	297
Total C		69.32		42.92	3,576
Heating Value, LHV Btu/lb (Btu/SCF)			19,080 (900)		

Source: DRI estimates based on data from Rio Blanco Oil Shale Co., February 1981.

TABLE 4.2-17. COMPOSITION OF DRIED FUEL GAS
(Stream 57)

Component	MWt	Mass %	Mole %	Mass Flow, 10^3 lb/hr	Flow, lb-moles/hr
H_2	2	4.68	41.82	2.89	1,447.3
CO	28	6.72	4.29	4.16	148.4
CO_2	44	1.00	0.41	0.62	14.1
N_2	28	6.72	4.29	4.16	148.4
NH_3	17	0.0028	0.0029	0.0017	0.1
H_2S	34	0.0006	0.0003	0.0003	0.01
CH_4	16	17.84	19.92	11.03	689.2
C_2H_4	28	11.53	7.35	7.13	254.5
C_2H_6	30	11.84	7.05	7.32	243.9
C_3H_6	42	14.77	6.28	9.13	217.3
C_3H_8	44	7.92	3.22	4.90	111.3
C_4H_8	56	12.97	4.14	8.01	143.1
C_4H_{10}	58	3.98	1.23	2.46	42.4
H_2O	18	0.01	0.01	0.0090	0.5
TOTAL		99.98	100.01	61.82	3,460.5
MWt			17.86		
Total H (excluding H_2O)		19.24		11.90	11,896
Total O (excluding H_2O)		4.58		2.83	177
Total N		6.73		4.16	297
Total C		69.41		42.92	3,576
Heating Value, LHV Btu/lb (Btu/SCF)			19,080 (900)		

Source: DRI estimates based on data from Rio Blanco Oil Shale Co., February 1981.

TABLE 4.2-18. MATERIAL BALANCE AROUND STRETFORD UNIT
(Streams 58, 61, 63, 64, 66)

Component	MWt	Acid Gases From DEA Regeneration (Stream 58) lb/hr	Acid Gases After Stretford (Stream 63) lb/hr	Stripping Air (Stream 61) lb/hr	Stretford Oxidizer Vent Gas (Stream 64) lb/hr	Stretford Sulfur (Stream 66) lb/hr (LTPSD)
H_2S	34	696	1.3	--	--	--
CO_2	44	55,344	52,023.0	--	3,321	--
N_2	28	--	--	16,117	16,117	--
O_2	32	--	--	4,633	4,285	--
H_2O	18	1,033	956.0	126	597	--
Sulfur (S_8)	256	--	--	--	--	695 (7.56)
TOTAL		57,073	52,980.3	20,876	24,320	695

Source: SWEC estimates based on information from Peabody Process Systems, Inc., February 1981.

levels of pollutants emitted, different flue gas compositions have not been calculated.

TABLE 4.2-19. COMPOSITION OF FLUE GAS FROM THE LIFT PIPES
(Stream 31)

Component	MWt	Mass %	Mole %	Mass Flow,* 10^3 lb/hr	(10^3 SCFM)
N_2	28	57.91	57.01	4,170.90	(942.13)
O_2	32	4.90	4.22	353.12	(69.79)
CO_2	44	20.04	12.56	1,443.63	(207.51)
H_2O	18	16.99	26.02	1,224.00	(430.08)
SO_2	64	0.0069	0.0030	0.50	(0.05)
CO	28	0.0092	0.0090	0.66	(0.15)
NOx	31.08	0.0339	0.0300	2.44	(0.50)
HC	16	0.0869	0.15	6.26	(2.47)
TPM	--	0.0154	--	1.11	--
TOTAL		99.99	100.00	7,202.62	(1,652.68)
MWt		27.56			
Total H (excluding H_2O)		0.02		1.57	
Total O (excluding H_2O)		19.51		1,405.00	
Total N		57.92		4,172.00	
Total C		5.54		398.70	
Total S		0.0035		0.25	

* SO_2, NOx, and CO assumed to be 30, 300, and 90 ppmv, respectively, in the flue gas. Particulate matter estimated to be 1,107 lb/hr. Includes steam from the quencher.

Source: DRI estimates based on data from Rio Blanco Oil Shale Co., February 1981.

4.2.7 Gas Liquor

The majority of the gas liquor (stream 41) is obtained as a result of the moisture condensation in the third tower. An additional amount of the liquor is produced as the compression condensate (stream 49) during the retort gas compression. These streams are combined to form the feed to the ammonia recovery plant in which anhydrous ammonia is recovered as a by-product. The compositions of the gas liquor and compression condensate are given in Table 4.2-20, while Table 4.2-21 presents the material balance around the ammonia unit. NaOH is added to release ammonia from ammonium sulfite.

4.2.8 Mine Water

Two aquifers are intercepted during the open pit mining. The composition of the aquifer waters, along with the values adopted in this manual, is presented in Table 4.2-22. Excess mine water, after process needs, is held in an aeration pond to reduce the alkalinity and to oxidize/consume organic matter. The composition of water ready for surface discharge is presented in Table 4.2-23.

4.3 POLLUTANT CROSS-REFERENCE TABLES

Tables 4.3-1 through 4.3-3 list some pollutants of concern, by medium, and provide a cross-reference to the numbered streams in this manual. Many of these pollutants are trace constituents, and measurements to identify or quantify them in oil shale processing related streams have never been made. Those pollutants which have been identified in the plant streams are cross-referenced to the detailed composition tables. Engineering judgment was used in identifying other probable pollutants. The entry for "unknown" (U) indicates that no testing has been done and the presence of the pollutant is unlikely. Judgment was also used in specifying the pollutants which definitely should not be present.

TABLE 4.2-20. COMPOSITION OF TOTAL FEED TO AMMONIA RECOVERY UNIT
(Streams 41, 49)

Component	Gas Liquor (Stream 41)		Compression Condensate (Stream 49)		Total Feed to Stripper	
	Mass %	lb/hr (gpm)	Mass %	lb/hr (gpm)	Mass %	lb/hr (gpm)
NH_3	0.59	1,758	0.45	17	0.59	1,775
CO_2	0.76	2,275	0.53	20	0.76	2,295
$(NH_4)_2SO_3$	0.13	400	0.76	29	0.14	429
Strippable Organics	0.02	66	--	--	0.02	66
Nonstrippable Organics	0.06	170	--	--	0.06	170
H_2O	98.43	293,000 (586)	98.26	3,735 (8)	98.43	296,735 (594)
TOTAL	99.99	297,669	100.00	3,801	100.00	301,470

Source: WPA estimates based on information from Rio Blanco Oil Shale Co., February 1981.

TABLE 4.2-21. MATERIAL BALANCE AROUND AMMONIA RECOVERY UNIT

(Streams 70, 71, 72)

Component	Feed to Stripper Column		Stripped Wastewater (Stream 70)		Ammonia Recovery Overhead Vapors (Stream 72)	Ammonia Product (Stream 71)
	Mass %	lb/hr (gpm)	Mass %	lb/hr (gpm)	lb/hr	lb/hr (TPSD)
NH_3	0.59	1,775	0.0014	4	14	1,883 (22.6)
CO_2	0.76	2,295	--	--	2,295	--
$(NH_4)_2SO_3$	0.14	429	--	--	--	--
Organics	0.08	236	0.06	170	66	--
NaOH*	0.10	301	0.0018	5	--	--
Na_2SO_3	--	--	0.17	466	--	--
H_2O	98.33	296,735 (594)	99.77	279,076 (558)	17,792	--
TOTAL	100.00	301,771	100.00	279,721	20,167	1,883

* NaOH is added to the steam stripping column to elevate pH and release fixed ammonia.

Source: WPA estimates.

TABLE 4.2-22. GROUNDWATER QUALITY OF LEASE TRACT C-a

(Stream 4)

Parameter (mg/l unless otherwise specified)	Lower Aquifer Min.	Lower Aquifer Max.	Lower Aquifer Avg.[a]	Upper Aquifer Min.	Upper Aquifer Max.	Upper Aquifer Avg.[a]	Values Adopted for Case Study[b] (Stream 4)
Alkalinity	52	4,500	674	70	2,390	408	560
Aluminum	<0.01	1.0	0.24	<0.01	1.0	0.15	0.2
Ammonia as NH_4	0.02	9.6	0.59	<0.1	1.8	0.3	0.47
Arsenic	<0.01	0.03	0.01	<0.01	0.05	0.010	0.01
Barium	<0.1	<1	<0.97	<0.1	<1	<0.98	<0.98
Beryllium	<0.1	<0.1	--[c]	<0.01	<0.01	<0.01	<0.01
Bicarbonate	260	3,310	842	85	2,760	482	687
Biochemical Oxygen Demand	--	--	--	--	--	--	--
Boron	0.01	5.7	0.84	0.01	4.8	0.33	0.62
Bromide	<0.02	<0.1	<0.05	<0.02	0.5	0.07	<0.05
Cadmium	<0.001	0.03	0.0099	<0.001	0.1	0.0096	0.0098
Calcium	0.80	98	8.8	4	260	34.8	20
Carbon, Dissolved Organic	3.0	73	10.5	1.0	50	8.5	9.6
Carbonate	<0.1	710	68.8	<0.1	335	0.88	40
Chloride	<0.1	160.0	21.7	<0.1	87	12.0	17.5
Chromium	<0.01	<0.05	<0.011	<0.01	<0.05	<0.012	<0.011
Chemical Oxygen Demand	<0.1	92	12.9	<0.1	400	17.7	15
Coliform, Fecal (col/100 ml)	--	--	--	--	--	--	--
Coliform, Total (col/100 ml)	--	--	--	--	--	--	--
Conductivity, μmho/cm	845	5,180	1,459	810	4,200	1,267	1,554
Copper	<0.01	0.3	0.088	<0.01	0.8	0.074	0.082
Cyanide	<0.01	0.08	0.01	<0.01	0.1	0.01	0.01
Fluoride	0.3	85	14.69	0.1	60	0.41	8.5
Hardness, as $CaCO_3$	20	630	110	32	1,110	328	204
Iron	<0.05	16.2	0.78	<0.005	36	5.02	2.6
Kjeldahl Nitrogen	--	--	--	--	--	--	--
Lead	0.003	26	0.21	0.002	3.8	0.17	0.19
Lithium	0.1	0.6	0.13	0.1	1.0	0.13	0.13
Magnesium	1.9	105	20	3.5	200	52	34
Manganese	0.05	0.8	0.075	<0.01	1.7	0.13	0.1
Methylene Blue Active Substances	--	--	--	--	--	--	--
Mercury	<0.001	<0.01	0.0024	<0.001	0.045	0.0036	0.0029
Molybdenum	<0.05	0.2	0.1	<0.05	1.8	0.13	0.11
Nickel	<0.001	<0.1	<0.023	<0.001	0.2	0.019	<0.023
Nitrate as NO_3	<0.1	2	0.1	<0.01	7	0.21	0.147
Nitrite as N	<0.20	0.60	0.07	<0.02	0.6	0.07	0.07
Nitrogen, Ammonia	0.40	3.2	1.18	0.20	1.8	0.51	0.89
pH (units)	6.0	8.9	7.16	6.0	8.8	6.78	7.00
Phenols	<0.001	1.0	0.0024	<0.001	0.17	0.0025	0.0025
Phosphate, Dissolved as PO_4	--	--	--	--	--	--	--

(Continued)

TABLE 4.2-22 (cont.)

Parameter (mg/l unless otherwise specified)	Lower Aquifer Min.	Lower Aquifer Max.	Lower Aquifer Avg.[a]	Upper Aquifer Min.	Upper Aquifer Max.	Upper Aquifer Avg.[a]	Values Adopted for Case Study[b] (Stream 4)
Phosphate, Ortho	<0.10	1.0	0.11	0.09	1.0	0.10	0.11
Potassium	<1.0	15.0	2.64	<1.0	11.0	2.19	2.44
Radioactivity, (pc/l)							
Gross alpha	0.1	30.0	3.31	0.1	29.0	3.48	3.38
Radium 226	0.09	0.9	0.31	0.1	0.8	0.17	0.25
Gross beta	2.0	830.0	21.4	2.0	73.0	12.7	17.6
Selenium	<0.01	<0.1	<0.010	<0.001	<0.01	<0.0098	<0.01
Silica as SiO_2	<0.1	60.0	10.1	<0.1	58	25.6	20.3
Silver	<0.001	0.1	0.0089	0.001	0.1	0.012	0.01
Sodium	155	1,560	397	92	1,170	212.0	317
Solids, Dissolved	540	3,640	1,075	530	2,850	905	1,002
Strontium	0.2	3.5	0.68	0.1	10.5	2.89	1.62
Sulfate	<4	580	112	<4	900	325	204
Sulfide	<0.01	6.50	0.56	0.03	49	0.63	0.59
Temperature (°F)	46.9	75.2	58.5	46.4	68.9	54.0	56.6
Vanadium	<0.05	<0.05	--	<0.05	<0.05	--	--
Zinc	0.02	68.0	0.24	0.01	15.0	0.26	0.25

[a] Arithmetic mean for pH and temperature, geometric mean for all other parameters.

[b] Based on 43% and 57% of upper and lower aquifer water production, respectively.

[c] Dashes (--) indicate data not reported.

Source: Gulf Oil Corp. and Standard Oil Co. (Indiana), May 1977.

TABLE 4.2-23. COMPOSITION OF EXCESS MINE WATER BEFORE AND AFTER AERATION
(Streams 75, 76)

Parameter, mg/l	Raw Mine Water* (Stream 75)	Treated Mine Water (Stream 76)
Alkalinity, as $CaCO_3$	560	500
Aluminum	0.2	0.2
Ammonia, Total	0.89	0.67
Arsenic	0.01	0.01
Boron	0.62	0.62
Calcium	20	20
Chloride	18	18
Chromium	<0.01	<0.01
COD	15	12
Cyanide	0.01	0.01
Fluoride	8.5	8.5
Lead	0.2	0.2
Mercury	0.003	0.003
pH (units)	7.0	~7
Phenols	0.0025	0.0025
Silica	20	20
Sodium	320	320
TDS	1,000	1,000
Sulfate	206	206
Sulfide	0.6	0.6
Flow Rate, gpm	8,330	8,330

* Based on data in Table 4.2-22, assuming mine water is 43% from upper and 57% from lower aquifer.

Source: WPA estimates based on data from Gulf Oil Corp. and Standard Oil Co. (Indiana), May 1977.

TABLE 4.3-1. POLLUTANT CROSS-REFERENCE FOR GASEOUS STREAMS

Stream No.	Table No.	Pollutants															
		CO	NO_x	SO_x	TPM	O_3	As	Be	Hg	Fluorides	H_2SO_4 Mist	H_2S	Pb	Total Reduced Sulfur	Vinyl Chloride	Polynuclear Organic Matter	Asbestos
5*		N	N	N	Y	N	Y	Y	Y	Y	N	N	Y	P	N	P	N
6*		N	N	N	Y	N	Y	Y	Y	Y	N	N	Y	P	N	P	N
7*		N	N	N	Y	N	Y	Y	Y	Y	N	N	Y	P	N	P	N
8*		N	N	N	Y	N	Y	Y	Y	Y	N	N	Y	P	N	P	N
9*		N	N	N	Y	N	Y	Y	Y	Y	N	N	Y	P	N	P	N
10*		N	N	N	Y	N	Y	Y	Y	Y	N	N	Y	P	N	P	N
11*		N	N	N	Y	N	Y	Y	Y	Y	N	N	Y	P	N	P	N
12*		N	N	N	Y	N	Y	Y	Y	Y	N	N	Y	P	N	P	N
13*		N	N	N	Y	N	Y	Y	Y	Y	N	N	Y	P	N	P	N
14*		N	N	N	Y	N	Y	Y	Y	Y	N	N	Y	P	N	P	N
15*		N	N	N	Y	N	Y	Y	Y	Y	N	N	Y	P	N	P	N
16*		N	N	N	Y	N	Y	Y	Y	Y	N	N	Y	P	N	P	N
17*		N	N	N	Y	N	Y	Y	Y	Y	N	N	Y	P	N	P	N
18*		N	N	N	Y	N	Y	Y	Y	Y	N	N	Y	P	N	P	N
19*		N	N	N	Y	N	Y	Y	Y	Y	N	N	Y	P	N	P	N
20*		N	N	N	Y	N	Y	Y	Y	Y	N	N	Y	P	N	P	N
21*		N	N	N	Y	N	Y	Y	Y	Y	N	N	Y	P	N	P	N
23*		N	N	N	Y	N	Y	Y	Y	Y	N	N	Y	P	N	P	N
24*		N	N	N	Y	N	Y	Y	Y	Y	N	N	Y	P	N	P	N
25		U	U	U	U	U	U	U	U	U	N	N	N	N	N	N	N
26	4.2-9	Y	Y	Y	P	N	P	U	U	U	U	N	Y	U	Y	N	N
27		N	N	N	N	N	N	N	N	N	N	N	N	N	N	N	N
30		P	P	P	Y	N	Y	Y	Y	Y	U	U	Y	P	N	P	N
31*	4.2-19	Y	Y	Y	Y	N	N	N	N	N	P	N	N	N	N	P	N
32*		N	N	N	Y	N	Y	Y	Y	Y	N	N	Y	P	N	P	N
33	4.2-11	Y	Y	Y	P	N	P	N	U	U	N	Y	U	Y	N	N	N
34	4.2-12	Y	Y	Y	P	N	P	N	U	U	N	Y	U	Y	N	N	N
35		Y	Y	Y	N	N	N	N	N	N	N	U	N	U	U	N	N
44*		Y	Y	Y	P	N	P	N	U	U	N	Y	U	Y	N	N	N
45	4.2-13	Y	N	Y	N	N	N	N	N	N	N	Y	N	Y	N	N	N

(Continued)

TABLE 4.3-1 (cont.)

Stream No.	Table No.	Pollutants															
		CO	NO_x	SO_x	TPM	O_3	As	Be	Hg	Fluorides	H_2SO_4 Mist	H_2S	Pb	Total Reduced Sulfur	Vinyl Chloride	Polynuclear Organic Matter	Asbestos
47*		Y	Y	Y	U	N	N	N	N	N	N	U	U	U	N	N	N
48	4.2-15	Y	N	Y	N	N	N	N	N	N	N	Y	N	Y	N	N	N
50		N	N	N	N	N	N	N	N	N	N	P	N	P	N	N	N
51		N	N	N	N	N	N	N	N	N	N	P	N	P	N	N	N
52		N	N	N	N	N	N	N	N	N	N	P	N	P	N	N	N
53		N	N	N	N	N	N	N	N	N	N	P	N	P	N	N	N
56	4.2-16	U	P	P	N	N	N	N	N	N	U	N	N	U	N	N	N
57	4.2-17	N	U	U	N	U	N	N	N	N	N	U	N	U	N	N	N
58	4.2-18	P	P	P	U	N	N	N	N	N	N	U	N	P	U	U	N
60*		P	P	P	N	N	N	N	N	N	P	P	N	U	U	U	N
61	4.2-18	P	U	U	N	U	N	N	N	N	N	U	N	U	N	N	N
63*	4.2-18	P	P	P	N	U	N	N	N	N	N	U	N	U	N	N	N
64	4.2-18	P	P	P	N	U	N	N	N	N	N	U	N	U	N	N	N
72	4.2-21	N	N	N	N	N	N	N	N	N	N	Y	N	Y	N	N	N
73*		N	N	N	Y	N	Y	Y	Y	Y	N	N	Y	P	N	P	N
74*		N	N	N	Y	N	Y	Y	Y	Y	N	N	Y	P	N	P	N
100*		N	N	N	N	N	N	N	N	N	N	N	N	N	N	N	N
107*		N	N	N	N	N	N	N	N	N	N	N	N	N	N	N	N
110*		N	N	N	N	N	N	N	N	N	N	N	N	N	N	N	N
112*		N	U	P	P	N	N	N	N	N	N	N	N	N	N	N	N

* Indicates streams that come into contact with the environment.

Key:
Y = Present
N = Not Present
P = Probably Present
U = Presence Unknown

Source: DRI estimates.

TABLE 4.3-2. POLLUTANT CROSS-REFERENCE FOR LIQUID STREAMS

Stream No.	Table No.	Pollutants																		
		Al	As	B	Cd	Cr	Co	F	Hg	Ni	Pb	Ra	V	Zn	NH_3	COD	TOC	Phenols	TDS	TSS
4*	4.2-22	P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	Y	Y	Y	Y
28*		P	P	P	P	P	P	P	P	P	P	P	P	P	P	P	P	P	Y	P
36	4.2-10	N	U	N	N	N	N	N	U	N	N	N	N	N	P	Y	Y	Y	P	P
37		N	U	N	N	N	N	N	U	N	N	N	N	N	P	Y	Y	Y	P	P
38	4.2-10	N	U	N	N	N	N	N	U	N	N	N	N	N	P	Y	Y	Y	P	P
39		N	U	N	N	N	N	N	U	N	N	N	N	N	P	Y	Y	Y	P	P
40		N	U	N	N	N	N	N	U	N	N	N	N	N	P	Y	Y	Y	P	P
41	4.2-20	N	U	N	N	N	N	N	U	N	N	N	N	N	Y	Y	Y	Y	Y	P
42	4.2-10	N	U	N	N	N	N	N	U	N	N	N	N	N	P	Y	Y	Y	P	Y
46	4.2-14	N	U	N	N	N	N	N	U	N	N	N	N	N	P	Y	Y	P	N	N
49	4.2-20	N	U	N	N	N	N	N	U	N	N	N	N	N	P	Y	Y	P	N	N
54		N	N	N	N	N	N	N	N	N	N	N	N	N	P	Y	Y	N	N	N
55		N	N	N	N	N	N	N	N	N	N	N	N	N	N	Y	Y	N	N	N
59*		N	N	N	N	N	N	N	N	N	N	N	N	N	P	Y	Y	N	P	P
62		N	N	N	N	N	N	N	N	N	N	N	Y	N	N	N	N	N	Y	N
65		N	N	N	N	N	N	N	N	N	N	N	Y	N	P	Y	Y	N	Y	P
66	4.2-18	N	N	N	N	N	N	N	N	N	N	N	P	N	N	N	N	N	N	N
67		N	N	N	N	N	N	N	N	N	N	N	N	N	N	N	N	N	N	N
69		N	N	N	N	N	N	N	N	N	N	N	N	N	N	N	P	N	N	N
70*	4.2-21	N	U	N	N	N	N	N	U	N	N	N	N	N	Y	Y	Y	Y	Y	N
71	4.2-21	N	N	N	N	N	N	N	N	N	N	N	N	N	Y	N	N	N	N	N
75*	4.2-23	P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	Y	Y	Y	Y
76*	4.2-23	P	P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	Y	Y	Y
77		P	P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	N	Y	N
78*		P	P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	Y	Y	Y
79		P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	Y	Y	Y	N
80		P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	Y	Y	Y	Y
81		N	N	N	N	N	N	N	N	N	N	N	N	N	N	N	P	N	N	N
82		P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	Y	Y	Y	N
83		P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	Y	Y	Y	N
84		N	N	N	N	N	N	N	N	N	N	N	N	N	N	N	P	N	N	N
85		N	N	N	N	N	N	N	N	N	N	N	N	N	N	N	P	N	N	N
86		P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	Y	Y	Y	N

(Continued)

TABLE 4.3-2 (cont.)

Stream No.	Table No.	Pollutants																		
		Al	As	B	Cd	Cr	Co	F	Hg	Ni	Pb	Ra	V	Zn	NH_3	COD	TOC	Phenols	TDS	TSS
87		P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	Y	Y	Y	N
88*		P	P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	P	Y	Y
89		P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	Y	Y	Y	N
90*		P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	Y	Y	Y	Y
91*		P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	P	P	Y	N
92*	4.2-3, 4.2-4	P	P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	P	Y	Y
93*		P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	Y	Y	Y	Y
94		P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	Y	Y	Y	Y
95*		P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	Y	Y	Y	Y
96*		P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	P	P	Y	Y
97		P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	P	P	Y	N
98		P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	P	P	Y	N
99*		P	P	P	P	P	P	P	P	P	P	P	P	P	P	P	P	P	Y	N
101*		P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	P	P	Y	Y
102*		P	P	P	P	P	P	P	P	P	P	P	P	P	N	N	N	N	Y	N
103*		P	P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	Y	Y	Y
104*		P	P	P	P	P	P	P	P	P	P	P	P	P	N	Y	Y	P	Y	Y
105*		P	P	P	P	P	P	P	P	P	P	P	P	P	N	Y	Y	P	Y	Y
106*		N	N	N	N	N	N	N	N	N	N	N	N	N	N	N	N	N	N	N
108*		P	P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	Y	Y	Y
109*		P	P	P	P	P	P	P	P	P	P	P	P	P	Y	Y	Y	P	Y	N
111*		P	P	P	P	P	P	P	P	P	P	P	P	P	N	P	P	P	Y	Y

* Indicates streams that come into contact with the environment.

Key: Y = Present
N = Not Present
P = Probably Present
U = Presence Unknown

Source: DRI estimates.

TABLE 4.3-3. POLLUTANT CROSS-REFERENCE FOR SOLID STREAMS

		Pollutants/Hazards															
Stream No.	Table No.	Ag	As	Ba	Cd	Cr	Hg	Pb	Se	Pesticides	Ignitability	Corrosivity	Reactivity, Explo.	Radio Activity	Phytotoxicity	Mutagenicity
1*	4.2-2	Y	Y	Y	Y	Y	Y	Y	Y	N	U	N	U	P	U	U
2*		Y	Y	Y	Y	Y	Y	Y	Y	N	U	N	U	P	U	U
3*		Y	Y	Y	Y	Y	Y	Y	Y	N	N	N	U	P	U	U
22*	4.2-2	Y	Y	Y	Y	Y	Y	Y	Y	N	U	N	U	P	U	U
29*	4.2-5, 4.2-6, 4.2-7	Y	Y	Y	Y	Y	Y	Y	Y	N	N	N	U	P	U	U
43		Y	Y	Y	Y	Y	Y	Y	Y	N	Y	N	U	P	U	U
68		N	N	N	N	N	N	N	N	N	N	Y	N	N	U	U

* Indicates streams that come into contact with the environment.

Key:
Y = Present
N = Not Present
P = Probably Present
U = Presence Unknown

Source: DRI estimates.

SECTION 5

POLLUTION CONTROL TECHNOLOGY

This section presents an inventory of pollution control technologies and discusses, in depth, some representative controls for each medium (air, water and solid waste). The inventory expands beyond describing the technologies that have been proposed for the Lurgi-Open Pit processes at Tract C-a. That is, it discusses alternate and additional technologies that provide varying levels of control. Although the inventory is quite extensive, other possibilities may exist and should not be excluded from consideration. Changes in the design of the plant complex, changes in the assumptions made (see Section 1.5), and/or improved data from future testing could lead to the selection of different controls.

Each subject area for control (e.g., particulate control) begins with an inventory of available technical approaches, or technologies. Promising new control technologies not yet applied commercially, even in related industries, are also included in the inventory but are not described in detail. Such new technologies may be applicable to the oil shale industry if they are sufficiently developed and tested in the future. The inventory is followed by a discussion of the most important considerations in selecting a control. Finally, a more detailed analysis of performance and cost is presented for the control technologies that have been considered by Rio Blanco Oil Shale Company (RBOSC) in conjunction with the Lurgi-Open Pit processes (see Sections 2 and 3 for a description of the case study which includes the proposed processes and technologies).

The detailed analysis seeks to estimate pollution control performance and cost. Performance estimates generally require no more than conceptual designs; however, the reliability of the performance estimates varies depending upon the application. The estimates should be highly reliable where a proven technology is applied to a conventional stream for which experience exists (e.g., flue gas desulfurization) but may be much less accurate for controls which require testing and which are applied to unconventional streams (e.g., biological oxidation). All performance levels are given for instantaneous control and reflect optimal operation, which may be higher than the average level of performance actually achieved. All cost estimates are in mid-1980 dollars and are taken to the level of detail believed to be necessary to achieve ±30% accuracy. This level of accuracy is based on the cost of equipment already built and operating in related industries.

5.1 AIR POLLUTION CONTROL

As in other industrial and oil shale operations, the Lurgi-Open Pit plant--from mining activities to final product storage and transfer--will generate particulate and gaseous component emissions. The primary air emissions are:

- Particulates, TPM
- Sulfur Dioxide, SO_2
- Nitrogen Oxides, NOx
- Carbon Monoxide, CO
- Hydrocarbons, HC.

This section describes the current, commercially available alternative systems for controlling the above primary pollutants. The following subsections provide inventories of control technologies for each of the air pollutants, a discussion of advantages and disadvantages, and important points to consider in selecting a particular technology. Performance, design, and cost data for the leading technologies examined are also presented.

5.1.1 Particulate Control

Particulate matter is generated during the mining, crushing, conveying, and processing of oil shale. Particulates are emitted from fugitive sources such as conveyor belts and from point sources such as flue gas stacks. Federal and State standards and regulations limit these particulate emissions because of their potentially hazardous effects on human health and the environment.

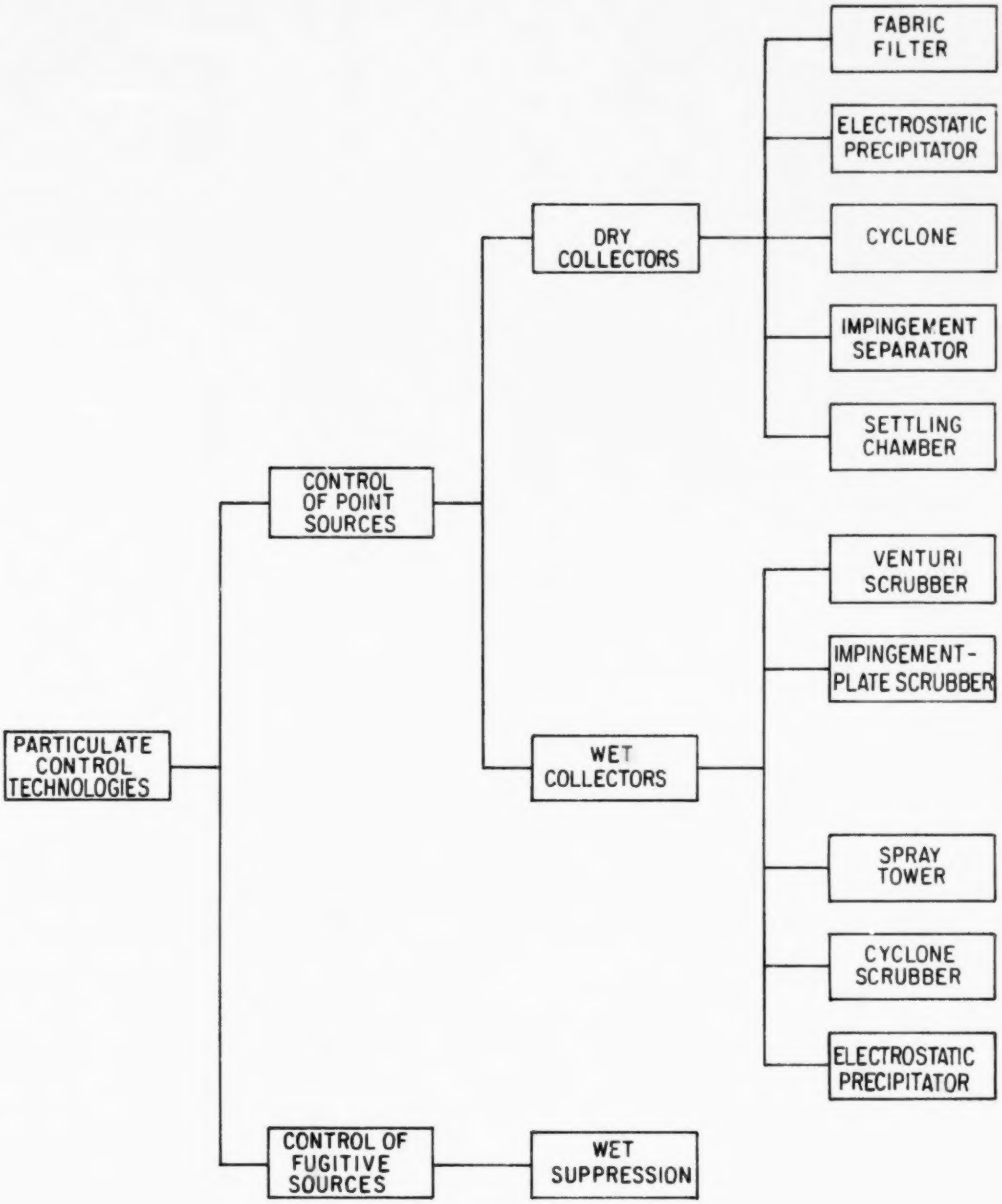
Inventory of Control Technologies--

As shown in Figure 5.1-1, particulate control can be divided into two general categories:

- Control of point sources
- Control of fugitive sources.

The particulate matter from a point source is confined within some equipment boundaries and is controlled by passing the dust-laden stream through a control device. Fugitive particulate matter is unconfined and is generally controlled by wet suppression techniques which are generally not as efficient as the point source control techniques. Table 5.1-1 presents a listing and review of particulate control technologies.

Control of point sources. There are two primary classes of particulate control equipment for point sources: dry and wet. Both classes offer processes that are feasible for particulate control in oil shale applications. Dry dust collectors can only be used with dry dusts. Sticky particulates tend to clog the dry collector and reduce its performance. In such cases, wet collectors are used.


PARTICULATE CONTROL TECHNOLOGIES
CONTROL OF POINT SOURCES
CONTROL OF FUGITIVE SOURCES
DRY COLLECTORS
WET COLLECTORS
WET SUPPRESSION
FABRIC FILTER
ELECTROSTATIC PRECIPITATOR
CYCLONE
IMPINGEMENT SEPARATOR
SETTLING CHAMBER
VENTURI SCRUBBER
IMPINGEMENT-PLATE SCRUBBER
SPRAY TOWER
CYCLONE SCRUBBER
ELECTROSTATIC PRECIPITATOR

SOURCE: SWEC

FIGURE 5.1-1 PARTICULATE CONTROL TECHNOLOGIES

TABLE 5.1-1. KEY FEATURES OF PARTICULATE CONTROL TECHNOLOGIES

Control Technology	Operating Principle	Performance	Development Status	Advantages	Disadvantages
DRY COLLECTORS					
Fabric Filter	The dust-laden gas passes through woven fabric or felt material which filters out the dust, allowing the gas to pass on. The filters are cleaned by mechanical shaking or reverse jet compressed air flow.	Removal efficiency is 99.7-99.9%. Operating temperature is limited to 600°F, depending on the fabric material, and the pressure drop is typically 4 in. H_2O.	Commercially proven.	High removal efficiency and low operating cost.	The fabric can be sensitive to gas humidity, velocity and temperature, as well as particulate characteristics.
Electrostatic Precipitator	Particles suspended in a gas are exposed to gas ions in an electrostatic field. These particles then become charged and migrate under the action of the field to collector plates.	Removal efficiency is 99-99.9%. Operating temperature is limited to 850°F, and the pressure drop is typically 1 in. H_2O.	Commercially proven.	High removal efficiency and a very low pressure drop.	High energy consumption. Sensitive to varying process conditions and particle properties.
Cyclone	The dust-laden gas enters a cylindrical or conical chamber tangentially at one or more points and leaves through a central opening. The dust particles, because of their inertia, will tend to move toward the outside separator wall and then into a receiver.	Removal efficiency is 50-90%. Operating temperature is limited to 1,000°F, and the pressure drop is typically 1-5 in. H_2O.	Commercially proven.	Low capital and operating cost. Good as a gas precleaner before a more efficient removal device.	Low removal efficiency for small particles.
Impingement Separator	The dust-laden gas impinges on a body, and the gas is deflected while the dust particle, by virtue of its greater inertia, collects on the surface of the body.	Removal efficiency is 0-80%. Operating temperature is limited to 1,000°F, and the pressure drop is typically 4 in. H_2O.	Commercially proven.	Low capital and operating cost. Good as a gas precleaner before a more efficient removal device.	Low removal efficiency.
Settling Chamber	The simplest type of dust collection equipment, consisting of a chamber in which the gas velocity is reduced to enable dust to settle out by the action of gravity.	Removal efficiency is 0-50%. Operating temperature is limited to 1,000°F, and the pressure drop is typically 0.1 in. H_2O.	Commercially proven.	Low operating cost and a low pressure drop.	Low removal efficiency and a very large space requirement.

(Continued)

TABLE 5.1-1 (cont.)

Control Technology	Operating Principle	Performance	Development Status	Advantages	Disadvantages
WET COLLECTORS					
Venturi Scrubber	Gas and liquid are passed concurrently through a venturi throat at 200 to 800 ft/sec.	Removal efficiency is 95-99%. Operating temperature is limited to 40-700°F, and the pressure drop is typically 1-50 in. H_2O.	Commercially proven.	High removal efficiency.	Efficiency drops for small particles.
Impingement-Plate Scrubber	The high velocity gas passes through a perforated tray with an impingement baffle above each perforation. The gas atomizes the liquid on the tray into droplets which then collect the dust.	Removal efficiency is 80-99%. Operating temperature is limited to 40-700°F, and the pressure drop is typically 1-20 in. H_2O.	Commercially proven.	High removal efficiency.	Efficiency drops for small particles.
Spray Tower	Liquid droplets produced by spray nozzles settle through rising gas stream and remove dust by impaction.	Removal efficiency is 50-80%. Operating temperature is limited to 40-700°F, and the pressure drop is typically 0.5 in. H_2O.	Commercially proven.	Low pressure drop and a low operating cost.	Low removal efficiency.
Cyclone Scrubber	Liquid is sprayed into the gas stream and removes the dust particles by inertial impaction.	Removal efficiency is 50-75%. Operating temperature is limited to 40-700°F, and the pressure drop is typically 2 in. H_2O.	Commercially proven.	Low pressure drop and a low operating cost.	Low removal efficiency.
Electrostatic Precipitator	SEE DRY COLLECTORS				
Wet Suppression	Fugitive dust generated in the crushing and handling of the oil shale is sprayed with a foam suppressant made from a water/surfactant mixture.	Removal efficiency is 95-99%. Operating temperature is limited to 40-200°F.	Commercially proven.	Low capital and operating cost and a high removal efficiency.	Used only for conveyor transfer points and crushing and grinding operations.

Source: SWEC based on information from Research and Education Association, 1980.

Dry particulate removal from gases can be accomplished by several methods (shown in Figure 5.1-1 and reviewed in Table 5.1-1). The baghouse (fabric filter) collector operates by passing the dust-laden gas through a fabric network that acts as a dust filter. Its removal efficiency is over 99% and operating costs are low compared to other approaches. The electrostatic precipitator, which also has a high removal efficiency, affords separation of dust by passing the dust-laden gas through an electrostatic field. The dust particles become charged and migrate to collector plates. Another technique is the cyclone separator which operates on the principle of centrifugal force. The dust-laden gas enters the chamber tangentially. The dust particles have a higher inertia than the gas, so they travel to the wall of the cylindrical or conical chamber and then into a receptacle. The removal efficiency for cyclones varies from 50 to 90 percent. In the impingement process, the dust-laden gas impinges on a body which collects the particles as the gas deflects around the body. The removal efficiency is 0 to 80 percent. The simplest mechanical separator is the settling chamber. The dust just settles to the bottom of the tank due to its heavier mass, resulting in a removal efficiency from 0 to 50 percent.

Wet collectors require mixing the dust-laden gas with an aqueous solution that captures and removes the dust particles from the gas stream. Examples of such equipment are shown in Figure 5.1-1 and described in Table 5.1-1. In the venturi scrubber, the gas and liquid pass through a throat at a high velocity, promoting collisions between the dust and liquid droplets. These units require a high pressure drop (~50 in. H_2O) and have removal efficiencies of greater than 90 percent. Another wet scrubber, the impingement-plate type, consists of a perforated tray with an impingement baffle located above each perforation. A liquid level maintained over the trays collects the dust as the gas passes through it. These units are similar to the venturi scrubber and have about the same removal efficiencies; however, the process requires a larger pressure drop across the plates. Another type of wet scrubber is the spray tower; it utilizes countercurrent spraying of liquid droplets to remove the dust particles by impaction at an efficiency of 50 to 80 percent. The operation of the wet cyclone scrubber is similar to that of the dry cyclone except liquid is introduced into the gas stream, removing the dust by inertial impaction. The removal efficiency of the wet cyclone is 50 to 75 percent. The wet electrostatic precipitator operates under the same principle and at about the same efficiency as the dry precipitator.

Control of fugitive sources. Wet dust suppression can be used for the containment of fugitive dust. This process consists of spraying the dust source with water or a foam suppressant which traps the dust and prevents it from becoming airborne. The foam sprays are relatively inexpensive, consume less water than pure water sprays, and are very effective. In a foam spray system, foam is produced by pumping a mixture of water and a surfactant through an air atomizing nozzle which produces small bubbles of approximately 100 to 200 microns. These wet bubbles are broken by contact with dust, coating the particle. The foam is only effective when applied directly to the source, such as on a conveyor, or to a falling stream of material, such as at a conveyor transfer point. Once the wetted dust agglomerates with other particles, it no longer becomes airborne at subsequent transfer points.

The suppression (or removal) efficiency is 95 to 99% in terms of the material contacted.

Paving the heavily traveled roads with a hard surface and providing vegetative covers for the disturbed areas are additional technologies for reducing fugitive dust emissions. These technologies are described in Section 5.3.

Particulate Control Technologies Analyzed--

A Lurgi-Open Pit plant will have several processing areas where particulate matter is generated and must be controlled. The particulates are produced either in the form of point source dust or as fugitive dust from various operations, such as raw shale mining, crushing, conveying and storage, Lurgi processed shale pyrolysis, solid waste handling and disposal, etc.

Dry dust is generated at point sources, such as the primary, secondary and tertiary crushers, closed conveyor transfer points, and enclosed ore storage areas. For these point sources, baghouses were examined for particulate removal due to their high removal efficiency, relatively low operating costs, and the dry nature of the particulate.

The air flow rates and particulate removal efficiencies for the baghouses were estimated from the stream flow rates and dust characteristics. Depending upon the type of dust, particle size distribution, and grain loading, the baghouses were designed with an air-to-cloth ratio of 5.9-6.7 to 1 (ACFM to sq. ft. of cloth), using Dacron HCE filter bags. The baghouses are equipped with adjustable pulse durations and cycle times for compressed air discharge through rotary airlock valves. When in multiples, the baghouses are manifolded to a common self-cleaning inlet, allowing part of any baghouse system to be shut down for repairs without taking the entire baghouse out of service.

Additional point source dust is generated in the Lurgi flue gas discharge system. The flue gas emerges from the waste heat recovery boilers at a temperature in excess of 425°F, which is well above the design temperature (275°F) for the polyester fabric baghouses. For this reason, conventional baghouses cannot be employed in this dust control application.

An electrostatic precipitator (ESP) was examined for the control of dust in the Lurgi flue gas. The performance of an ESP, unlike baghouses, is not affected adversely by the high temperature of the dust. On the contrary, within limits, this characteristic of the dust increases the efficiency of dust removal by ESP. The use of an ESP is in agreement with the plans presented by RBOSC, but design details were not given (Rio Blanco Oil Shale Co., February 1981). Information provided by Research Cottrell Corporation (an ESP vendor) was used in conjunction with Stone and Webster Engineering Corporation's (SWEC) in-house experience to estimate the design and cost of an ESP for the Lurgi flue gas dust control application.

Fugitive dust is generated from mining, raw and processed shale conveying, conveyor transfer points, baghouse dust discharge points to conveyor systems, truck loading and unloading, and disposal operations. These fugitive sources of particulates are controlled by water and foam spray suppression. This system is inexpensive and offers low water consumption and high removal efficiency.

Table 5.1-2 lists the design parameters for the particulate control technologies examined, Table 5.1-3 presents more design details for the baghouses, and Table 5.1-4 gives the design basis for the ESP. The capital, operating, and annual costs for tne particulate control equipment are presented in Table 5.1-5. Figures 5.1-2 and 5.1-3 present the cost curves for the baghouses and ESP, respectively. The curves have been derived specifically from the stream characteristics and design parameters used in this manual.

Other Particulate Control Technologies Analyzed--

In addition to the ESP, another technology was analyzed for the control of particulates from the Lurgi flue gas--a fiberglass fabric baghouse. This technology has not been proposed by RBOSC, but it is judged to be applicable to the flue gas.

As mentioned previously, the flue gas is at a fairly high temperature at the point of control; therefore, conventional polyester fabric baghouses cannot be used for dust control. The fiberglass reinforced fabric baghouses, on the other hand, have a much higher operating temperature limit (600°F). The operation of the latter type of baghouses is similar to that of conventional baghouses, and comparable dust removal efficiencies are obtained. Table 5.1-6 presents the design and cost details for the fiberglass baghouse analyzed for the Lurgi flue gas application. The cost curve for conventional baghouses (see Figure 5.1-2) can be used for fiberglass baghouses because, except for the fabric material, the two types of baghouses are quite similar.

Total Particulate Emissions--

The controlled particulates from the point as well as fugitive sources are summarized in Table 5.1-7, along with the type of control technology examined for each source. The uncontrolled emissions are also included in the table to give total particulate emissions from the commercial operation. Estimates for these emissions were based on information provided by the equipment vendors.

5.1.2 Sulfur Control

Processing of sulfur-containing fossil fuels will result in emissions of sulfur compounds, such as H_2S, COS, CS_2, RSH, etc., or their combustion product, SO_2. Federal and State standards limit sulfur emissions because of their potentially hazardous effects on human health and the environment.

Inventory of Control Technologies--

Two general categories of technologies are available for the control of sulfur emissions: (1) removal of sulfur compounds from flue gases

TABLE 5.1-2. PARTICULATE CONTROL EQUIPMENT AND DESIGN PARAMETERS

Stream Number	Control Description	Control Location	Number of Units	Flow Rate Each (ACFM)	Dust Load (lb/hr)	Removal Efficiency (%)	Total Particulate Emissions (lb/hr)
5	Baghouse	Primary Crushers (ore)	2	61,100	5,237.1	99.7	15.71
6	Baghouse	Primary Crushers (subore)	1	12,200	522.9	99.7	1.57
7	Baghouse	Primary Crushers (overburden)	1	63,800	2,734.3	99.7	8.20
8	Baghouse	Conveyor to Stockpile	2	36,300	124.5	99.7	0.37
9	Baghouse	Raw Shale Conveyor Transfer Points	3	40,500	208.3	99.7	0.62
10	Baghouse	Conveyor to Secondary Crushers	2	20,200	69.3	99.7	0.21
11	Baghouse	Secondary Crushers	8	69,800	23,931.4	99.7	71.79
12	Baghouse	Conveyor to Secondary Screens	2	20,200	69.3	99.7	0.21
13	Baghouse	Secondary Screens	8	69,800	23,931.4	99.7	71.79
14	Baghouse	Conveyor to Tertiary Crushers	2	20,200	69.3	99.7	0.21
15	Baghouse	Tertiary Crushers	9	69,800	26,922.9	99.7	80.77
16	Baghouse	Conveyor to Tertiary Screens	4	20,200	138.5	99.7	0.42
17	Baghouse	Tertiary Screens	9	69,800	26,922.9	99.7	80.77
18	Baghouse	Conveyor to Fine Ore Storage	2	20,200	69.3	99.7	0.21
19	Baghouse	Fine Ore Storage	1	28,800	1,234.3	99.7	3.70
20	Baghouse	Conveyor to Retort Feed Hoppers	2	20,200	69.3	99.7	0.21
21	Baghouse	Retort Feed Hoppers	4	53,100	9,102.9	99.7	27.31
23	Baghouse	Processed Shale Load-out Hoppers	3	21,500	2,764.3	99.7	8.29
31	Electrostatic Precipitator	Flue Gas Discharge System	13	293,700	1,106,554.8	99.9	1,106.6
32	Baghouse	Conveyor to Retorts	13	18,400	410.1	99.7	1.23
73	Baghouse	Processed Shale Conveyor Transfer Points	2	32,300	110.7	99.7	0.33
74	Water and Foam Sprays	Open Stockpiles, etc.	--	--	3,466.7	98.5	52.0

Source: SWEC estimates based on information from Gulf Oil Corp. and Standard Oil Co. (Indiana), March 1976, and Rio Blanco Oil Shale Co., February 1981.

TABLE 5.1-3. BAGHOUSE SPECIFICATIONS*

Stream Number	Control Location (No. of Units)	Flow Rate Each (ACFM)	No. of Bags	Net Cloth Area (ft^2)	Air-to-Cloth Ratio (ft^2/ACF)	Fan ΔP (in. H_2O)	Fan Motor (BHP)
5	Primary Crusher (ore) (2)	61,100	882	10,399	5.87/1	9.5	2X62
6	Primary Crusher (subore) (1)	12,200	176	2,076	5.87/1	9.5	2X12
7	Primary Crusher (overburden) (1)	63,800	921	10,858	5.87/1	9.5	2X64
8	Conveyor to Stockpile (2)	36,300	461	5,428	6.68/1	9.5	2X34
9	Raw Shale Conveyor Transfer Points (3)	40,500	514	6,057	6.68/1	9.5	2X38
10	Conveyor to Secondary Crushers (2)	20,200	256	3,021	6.68/1	9.5	2X19
11	Secondary Crushers (8)	69,800	1,008	11,879	5.87/1	9.5	2X70
12	Conveyor to Secondary Screens (2)	20,200	256	3,021	6.68/1	9.5	2X19
13	Secondary Screens (8)	69,800	1,008	11,879	5.87/1	9.5	2X70
14	Conveyor to Tertiary Crushers (2)	20,200	256	3,021	6.68/1	9.5	2X19
15	Tertiary Crushers (9)	69,800	1,008	11,879	5.87/1	9.5	2X70
16	Conveyor to First Set of Tertiary Screens (2)	20,200	256	3,021	6.68/1	9.5	2X19
16	Conveyor to Second Set of Tertiary Screens (2)	20,200	256	3,021	6.68/1	9.5	2X19
17	Tertiary Screens (9)	69,800	1,008	11,879	5.87/1	9.5	2X70
18	Conveyor to Fine Ore Storage (2)	20,200	256	3,021	6.68/1	9.5	2X19
19	Fine Ore Storage (1)	28,800	416	4,901	5.87/1	9.5	2X29
20	Conveyor to Retort Feed Hoppers (2)	20,200	256	3,021	6.68/1	9.5	2X19
21	Retort Feed Hoppers (4)	53,100	767	9,037	5.87/1	9.5	2X54
23	Processed Shale Load-out Hoppers (3)	21,500	310	3,659	5.87/1	9.5	2X22
32	Conveyor to Retorts (13)	18,400	234	2,752	6.68/1	9.5	2X17
73	Processed Shale Conveyor Transfer Points (2)	32,300	466	5,497	5.87/1	9.5	2X33

* Other items included in the estimate are site preparation and concrete foundations, ductwork (3/16" C.S. Plate), electrical, instrumentation, and duct and collector insulation. In addition, the fabric used is 16-oz. Dacron HCE.

Source: SWEC estimates based on information provided by North-Monson Co., August 11, 1980.

TABLE 5.1-4. MAJOR ITEMS IN ELECTROSTATIC PRECIPITATOR*

Capital Cost Items	Operating Cost Items
Chambers (13)	Electricity 4,907 kW
Collecting Plates	Maintenance
Transformer Rectifiers	
Fans and Motors	
Dampers and Ductwork	
Supports	
Handrailing and Grating	
Piping	
Concrete and Foundations	
Painting	
Insulation	
Instrumentation and Controls	
Discharge Electrodes	
Electrical	
Bins	
Discharge and Conveying System	
Rappers	

* Design basis: 293,700 ACFM/unit.

Source: SWEC estimates based on information provided by Research Cottrell Corp.

TABLE 5.1-5. COST OF PARTICULATE POLLUTION CONTROL

Stream Number	Control Description	Control Location	Number of Units	Flow Rate Each (ACFM)	Fixed Capital Cost ($000's)	Total Annual Operating Cost ($000's)	Total Annual Control Cost ($000's)*
5	Baghouse	Primary Crushers (ore)	2	61,100	987	63	250
6	Baghouse	Primary Crushers (subore)	1	12,200	105	6	26
7	Baghouse	Primary Crushers (overburden)	1	63,800	552	34	138
8	Baghouse	Conveyor to Stockpile	2	36,300	628	85	205
9	Baghouse	Raw Shale Conveyor Transfer Points	3	40,500	1,051	65	264
10	Baghouse	Conveyor to Secondary Crushers	2	20,200	348	22	88
11	Baghouse	Secondary Crushers	8	69,800	4,828	297	1,211
12	Baghouse	Conveyor to Secondary Screens	2	20,200	348	22	88
13	Baghouse	Secondary Screens	8	69,800	4,828	297	1,211
14	Baghouse	Conveyor to Tertiary Crushers	2	20,200	348	22	88
15	Baghouse	Tertiary Crushers	9	69,800	5,432	334	1,363
16	Baghouse	Conveyor to Tertiary Screens	4	20,200	696	44	176
17	Baghouse	Tertiary Screens	9	69,800	5,432	334	1,363
18	Baghouse	Conveyor to Fine Ore Storage	2	20,200	348	22	88
19	Baghouse	Fine Ore Storage	1	28,800	249	15	62
20	Baghouse	Conveyor to Retort Feed Hoppers	2	20,200	348	22	88
21	Baghouse	Retort Feed Hoppers	4	53,100	1,837	113	471

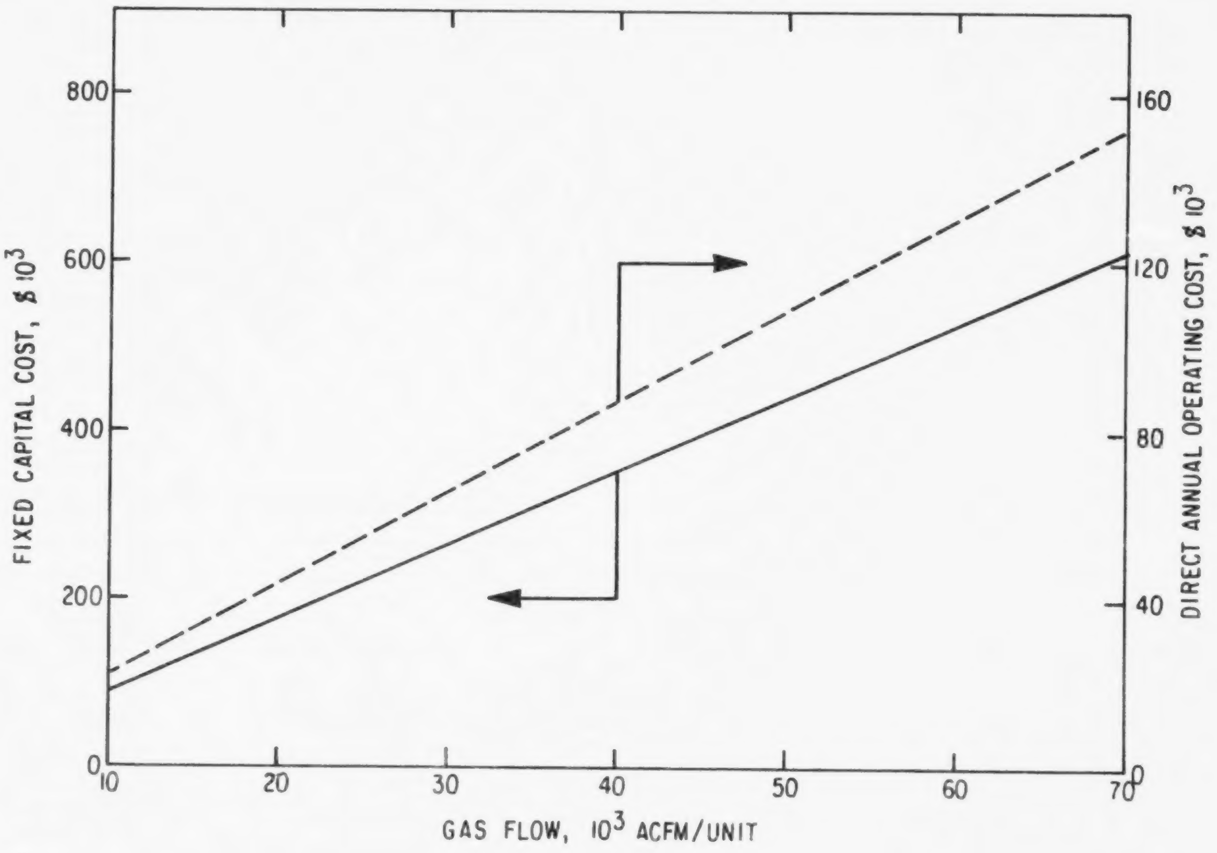
(Continued)

TABLE 5.1-5 (cont.)

Stream Number	Control Description	Control Location	Number of Units	Flow Rate Each (ACFM)	Fixed Capital Cost ($000's)	Total Annual Operating Cost ($000's)	Total Annual Control Cost ($000's)*
23	Baghouse	Processed Shale Load-out Hoppers	3	21,500	559	34	140
31	Electrostatic Precipitator	Flue Gas Discharge System	13	293,700	50,734	2,144	12,016
32	Baghouse	Conveyor to Retorts	13	18,400	2,059	127	528
73	Baghouse	Processed Shale Conveyor Transfer Points	2	32,300	559	34	140
74	Water and Foam Sprays	Open Stockpiles, etc.	--	--	909	1,456	1,650
TOTAL					83,185	5,592	21,654

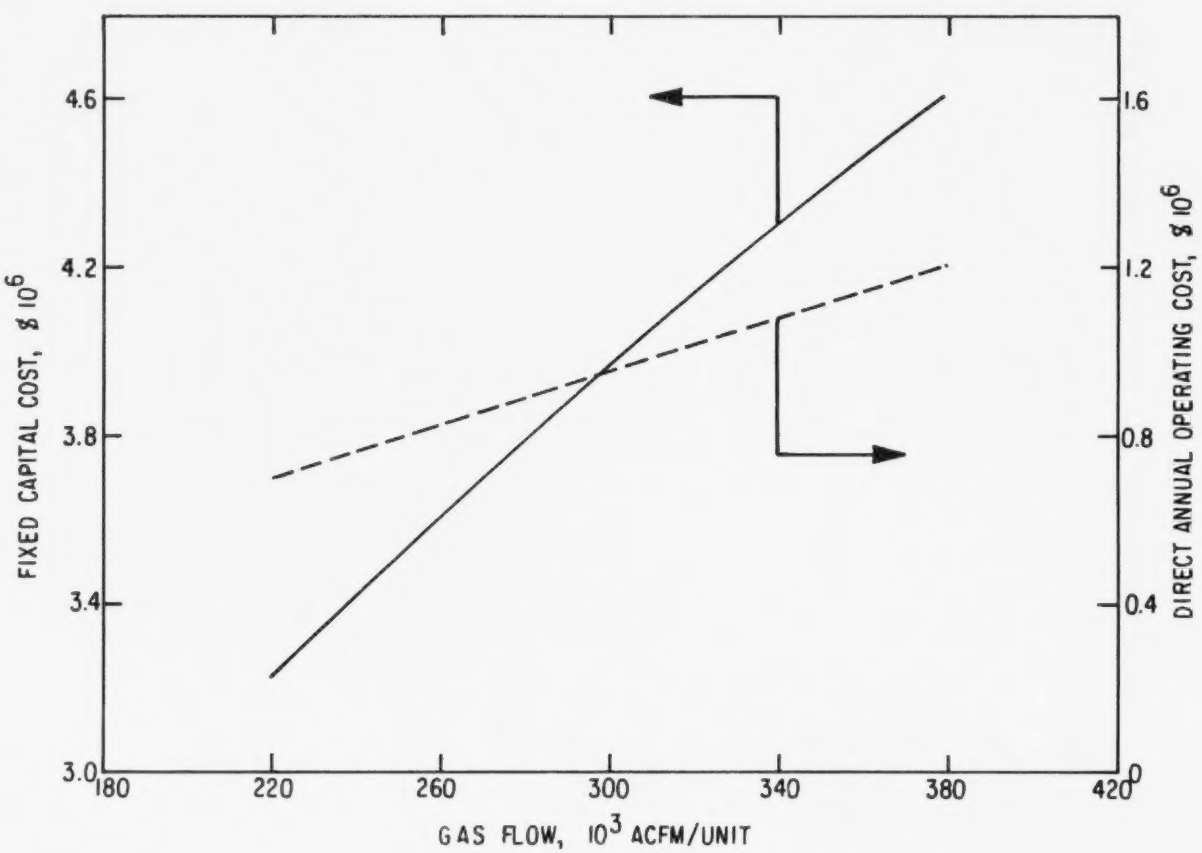
* See Section 6 for details on computation of the total annual control cost.

Source: DRI estimates based on information provided by SWEC.



SOURCE: DRI based on information provided by SWEC

FIGURE 5.1-2 COST OF PARTICULATE CONTROL WITH BAGHOUSES



SOURCE: DRI based on information provided by SWEC

FIGURE 5.1-3 COST OF PARTICULATE CONTROL WITH ELECTROSTATIC PRECIPITATORS

TABLE 5.1-6. DESIGN AND COST OF THE FIBERGLASS FABRIC BAGHOUSE

Item	Unit	Quantity
No. of Baghouses	--	13
Flow Rate (each)	ACFM	293,700
No. of Bags (each)	--	3,774
Net Cloth Area (each)	ft^2	44,462
Air-to-Cloth Ratio	ft^2/ACF	6.6
Fan, ΔP	in. H_2O	10.5
Fan Motor	BHP	2 x 315
Dust Loading	grains/ACF	33
	10^3 lb/hr	1,106.6
Dust Removal Efficiency	%	99.7
Outlet Dust Concentration	lb/hr	3,319.7
Fixed Capital Cost	$\$10^3$	33,015
Direct Annual Operating Cost	$\$10^3$	
Maintenance		644
Electricity		1,279
TOTAL		1,923

Source: SWEC estimates based on information provided by North-Monson Co., August 11, 1980.

TABLE 5.1-7. TOTAL PARTICULATE EMISSIONS FROM THE PLANT

Stream Number	Emission Source	Control Description	Particulate Emissions (lb/hr)
5	Primary Crushers (ore)	Baghouse	15.71
6	Primary Crushers (subore)	Baghouse	1.57
7	Primary Crushers (overburden)	Baghouse	8.20
8	Conveyor to Stockpile	Baghouse	0.37
9	Raw Shale Conveyor Transfer Points	Baghouse	0.62
10	Conveyor to Secondary Crushers	Baghouse	0.21
11	Secondary Crushers	Baghouse	71.79
12	Conveyor to Secondary Screens	Baghouse	0.21
13	Secondary Screens	Baghouse	71.79
14	Conveyor to Tertiary Crushers	Baghouse	0.21
15	Tertiary Crushers	Baghouse	80.77
16	Conveyor to Tertiary Screens	Baghouse	0.42
17	Tertiary Screens	Baghouse	80.77
18	Conveyor to Fine Ore Storage	Baghouse	0.21
19	Fine Ore Storage	Baghouse	3.70
20	Conveyor to Retort Feed Hoppers	Baghouse	0.21
21	Retort Feed Hoppers	Baghouse	27.31

(Continued)

TABLE 5.1-7 (cont.)

Stream Number	Emission Source	Control Description	Particulate Emissions (lb/hr)
23	Processed Shale Load-out Hoppers	Baghouse	8.29
24	Diesel Equipment	--	33.00
31	Flue Gas Discharge System	Electrostatic Precipitator	1,106.6
32	Conveyor to Retorts	Baghouse	1.23
73	Processed Shale Conveyor Transfer Points	Baghouse	0.33
74	Open Stockpiles, etc.	Water and Foam Sprays	52.00
TOTAL			1,565.52

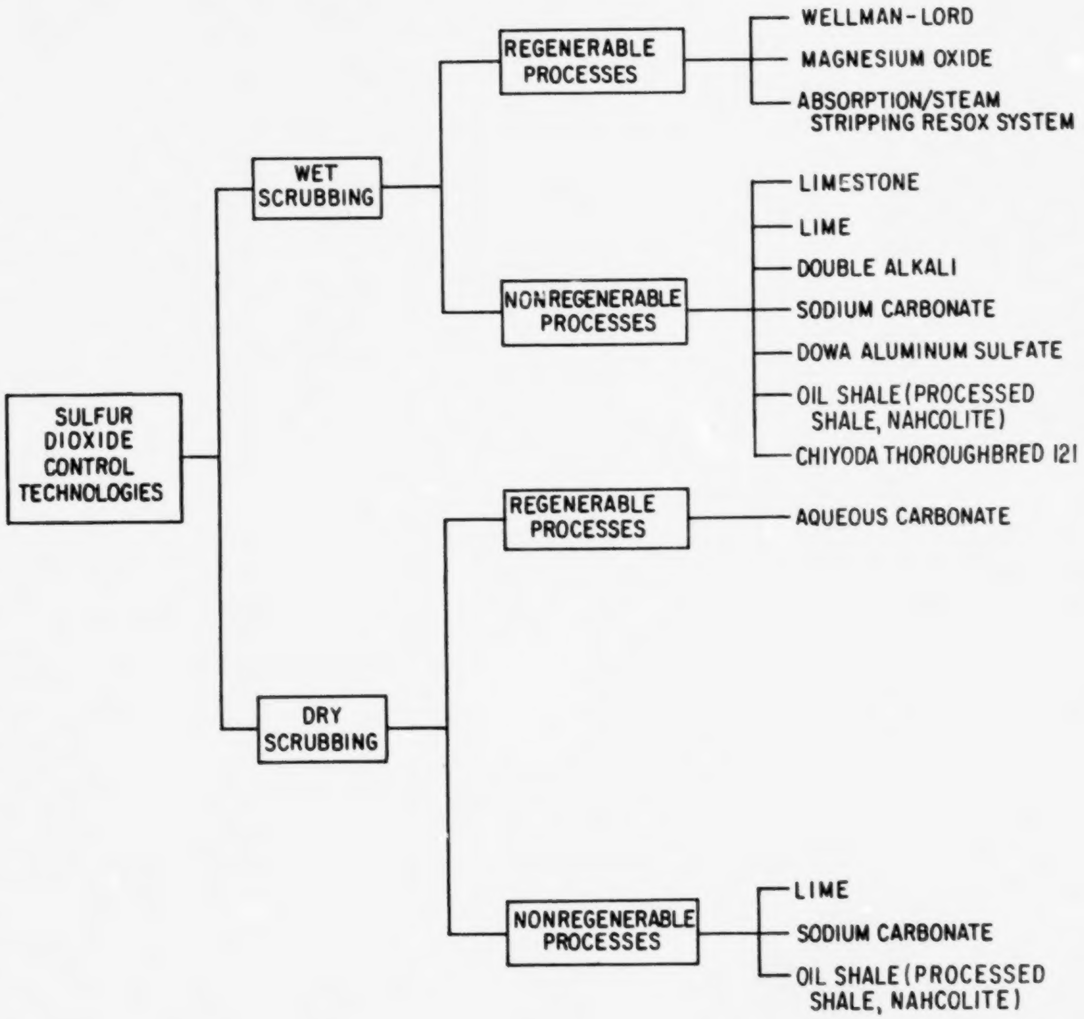
Source: SWEC estimates based on information from Gulf Oil Corp. and Standard Oil Co. (Indiana), March 1976, and Rio Blanco Oil Shale Co., February 1981.

after combustion (sulfur dioxide removal, or flue gas desulfurization) and (2) removal of sulfur compounds from gases prior to combustion (hydrogen sulfide removal). Several technologies in both categories offer recovery of sulfur in a useful form, while others chemically fix the sulfur compounds on a reagent which then requires disposal.

Sulfur dioxide control (flue gas desulfurization). Removal of sulfur compounds from flue gases--that is, flue gas desulfurization (FGD)--is based on the physical and chemical properties of SO_2 because fuel-based sulfur is usually converted to SO_2 upon combustion. Flue gas desulfurization can be divided into two categories:

- Wet scrubbing
- Dry scrubbing.

Wet scrubbing utilizes a solution or a slurry to absorb the SO_2. Dry scrubbing uses either a dry reagent bed or an atomized solution of an aqueous reagent at a high temperature to remove the SO_2. Both categories can be divided into regenerable and nonregenerable processes. The different types of SO_2 removal processes are shown in Figure 5.1-4, and Table 5.1-8 gives a brief description of each process.


SULFUR DIOXIDE CONTROL TECHNOLOGIES
WET SCRUBBING
REGENERABLE PROCESSES
WELLMAN-LORD
MAGNESIUM OXIDE
ABSORPTION/STEAM STRIPPING RESOX SYSTEM
NONREGENERABLE PROCESSES
LIMESTONE
LIME
DOUBLE ALKALI
SODIUM CARBONATE
DOWA ALUMINUM SULFATE
OIL SHALE (PROCESSED SHALE, NAHCOLITE)
CHIYODA THOROUGHBRED 121
DRY SCRUBBING
REGENERABLE PROCESSES
AQUEOUS CARBONATE
NONREGENERABLE PROCESSES
LIME
SODIUM CARBONATE
OIL SHALE (PROCESSED SHALE, NAHCOLITE)

SOURCE: SWEC

FIGURE 5.1-4 SULFUR DIOXIDE CONTROL TECHNOLOGIES

TABLE 5.1-8. KEY FEATURES OF SULFUR DIOXIDE CONTROL TECHNOLOGIES

Control Technology	Operating Principle	Performance*	Development Status	Advantages	Disadvantages
REGENERABLE WET SCRUBBING PROCESSES					
Wellman-Lord[a]	Absorbs SO_2 with a sodium sulfite/bisulfite solution. A bleed stream of the rich solution is sent to evaporators where SO_2 and water are driven off and the reagent is regenerated.	Reduces the outlet flue gas SO_2 concentration to 50 ppmv.	About 30 commercial units are in operation in the U.S. and Japan.	Produces a concentrated SO_2 stream which can be used to make salable sulfur or sulfuric acid.	Requires fuel for the solution evaporators.
Magnesium Oxide[a]	Absorbs SO_2 with a magnesium oxide slurry. A bleed stream of the spent slurry is dried and calcined to regenerate the magnesium oxide and produce a dilute SO_2 stream (10% SO_2).	Reduces the outlet flue gas SO_2 concentration to 50 ppmv.	Three demonstration plants have been tested in Japan (each about 100 MW size). Two commercial units are under construction in the U.S.	Produces an SO_2 stream suitable for manufacture of sulfuric acid.	Requires fuel for the $MgSO_3/MgSO_4$ dryer and calciner.
Absorption/ Steam Stripping Resox System[a]	An aqueous solution of a suitable reagent (e.g., sodium carbonate, citric acid) absorbs the SO_2, and the solution is regenerated by indirect steam heating to evolve a concentrated SO_2 stream. The SO_2 is then reacted with crushed coal in the Resox system to produce elemental sulfur.	Reduces the outlet flue gas SO_2 concentration to 50 ppmv.	The systems have been tested in separate demonstration plants. A demonstration plant for the combined system has been proposed.	Uses a simple absorption/ steam stripping system and produces salable elemental sulfur.	Has not been demonstrated as a combined system.
NONREGENERABLE WET SCRUBBING PROCESSES					
Limestone[a]	Absorbs SO_2 with a limestone slurry. A bleed stream of the slurry is partially dewatered and disposed of in a landfill.	Reduces the outlet flue gas SO_2 concentration to 50 ppmv.	Many commercial units in operation worldwide.	Low capital and operating cost. Simple and proven process with conventional process equipment.	Has a low operability factor due to scaling, erosion and corrosion. Sulfur is nonrecoverable.
Lime[a]	Absorbs SO_2 with a lime slurry. A bleed stream of the slurry is partially dewatered and disposed of in a landfill.	Reduces the outlet flue gas SO_2 concentration to 50 ppmv.	Many commercial units in operation worldwide.	Very similar to the limestone process and can potentially give a greater SO_2 removal efficiency than limestone.	Lime costs are rising rapidly because of higher energy costs. Sulfur is nonrecoverable.

(Continued)

TABLE 5.1-8 (cont.)

Control Technology	Operating Principle	Performance*	Development Status	Advantages	Disadvantages
Double Alkali[a]	Uses two alkaline solutions, sodium hydroxide and sodium sulfite, to convert SO_2 to sodium bisulfite. The spent solution is regenerated by lime addition. The precipitated solids are partially dewatered and disposed of in a landfill.	Reduces the outlet flue gas SO_2 concentration to ≤ 50 ppmv.	Nine commercial units in operation and three under construction in the U.S.	Low in capital and operating cost like the limestone system, but the use of a clear scrubbing solution reduces scaling, erosion, and corrosion in the scrubbing loop.	Requires soda ash (Na_2CO_3) makeup in addition to lime for precipitation. Soda ash is an expensive raw material. The sludge contains soluble and leachable sodium salts. Sulfur is nonrecoverable.
Sodium Carbonate[a]	Absorbs SO_2 with a sodium carbonate solution. A bleed stream of the spent solution is partially dewatered and disposed of in a landfill.	Reduces the outlet flue gas SO_2 concentration to 50 ppmv.	Four commercial units in operation worldwide.	Low capital cost. A very simple and reliable process.	Soda ash is an expensive raw material. Produces a sludge which is very difficult to dewater and dispose of. Sulfur is nonrecoverable.
Dowa Aluminum Sulfate[b]	Absorbs SO_2 with an acidic clear solution of basic aluminum sulfate. The spent solution is oxidized to aluminum sulfate. Limestone is added to the solution to regenerate basic aluminum sulfate and produce gypsum which is partially dewatered and disposed of in a landfill.	Reduces the outlet flue gas SO_2 concentration to 50 ppmv.	Not used commercially in the U.S.	Uses the same basic process design as the double alkali process and, therefore, has the same advantages. The process uses a clear scrubbing solution which reduces scaling, erosion, and corrosion in the scrubbing loop.	Requires dewatering and landfill disposal.
Oil Shale[b] (Processed Shale, Nahcolite)	Absorbs SO_2 with a shale slurry. A bleed stream of the slurry is partially dewatered and disposed of in a landfill.	Reduces the outlet flue gas SO_2 concentration to ~50 ppmv.	The process is only conceptual at this time and has not been tested on a pilot scale.	Low capital and operating cost. Also, an abundant supply of processed shale is available at the plant site. Nahcolite is plentiful in the Piceance Basin.	Has not yet been tested, even on a pilot plant scale. Sulfur is nonrecoverable.

(Continued)

TABLE 5.1-8 (cont.)

Control Technology	Operating Principle	Performance*	Development Status	Advantages	Disadvantages
Chiyoda Thoroughbred 121 (CT-121)[c]	The flue gas is first quenched to its saturation temperature and then sparged into a limestone slurry, generating a jet bubbling froth layer. The SO_2 in the flue gas is absorbed by the limestone slurry in the jet bubbling layer. The calcium sulfite formed by this reaction is oxidized to calcium sulfate (gypsum) by the introduction of air into the jet bubbling layer. A bleed stream of the waste slurry can be dewatered and landfilled as a recoverable resource or given away to local cement, fertilizer or wallboard industries.	Reduces the outlet flue gas SO_2 concentration to 50 ppmv.	The process has been tested on a demonstration size scale.	Absorbs SO_2 and oxidizes calcium sulfite to gypsum in one reactor vessel.	Has only been tested on a demonstration size scale. Sulfur is nonrecoverable.
REGENERABLE DRY SCRUBBING PROCESS					
Aqueous Carbonate[a]	Flue gas is contacted with an atomized solution of aqueous sodium carbonate in a spray dryer scrubber. The sodium carbonate absorbs the SO_2, is dried, and then collected in a baghouse or electrostatic precipitator (ESP). The dried product is sent to the reducer vessel where it is reacted with coal in a molten sodium carbonate and sodium sulfide solution to form sodium sulfide. The sodium sulfide is then reacted in another vessel with CO_2 in the off-gas from the reducer vessel to regenerate sodium carbonate and evolve hydrogen sulfide gas. The hydrogen sulfide gas is sent to a Claus unit to produce elemental sulfur.	Reduces the outlet flue gas SO_2 concentration to 75-100 ppmv.	The process steps have been tested individually on a pilot scale. An integrated demonstration size (100 MW) unit is currently under construction in the U.S.	Produces salable elemental sulfur using coal as a fuel instead of higher priced and less available oil and natural gas.	The hot molten carbonate solution used in the regeneration section of the process is very corrosive and will require very expensive, special construction materials.

(Continued)

TABLE 5.1-8 (cont.)

Control Technology	Operating Principle	Performance*	Development Status	Advantages	Disadvantages
NONREGENERABLE DRY SCRUBBING PROCESSES					
Lime[a]	Flue gas is contacted with an atomized lime slurry in a spray dryer scrubber. The lime absorbs the SO_2, is dried, and then collected in a baghouse or electrostatic precipitator (ESP).	Reduces the outlet flue gas SO_2 concentration to 100-150 ppmv.	Two commercial units are in operation in the U.S. Three additional units are currently under construction.	Since the flue gas is not saturated, slightly less makeup water is needed and less stack gas reheat is needed.	This system is usually only economically feasible where low sulfur fuel is burned because of the low reagent utilization rate. Very high removal efficiencies are also not usually possible because of the low reagent utilization rate. Sulfur is nonrecoverable.
Sodium Carbonate[a]	Flue gas is contacted with an atomized solution of aqueous sodium carbonate in a spray dryer scrubber. The sodium carbonate absorbs the SO_2, is dried, and then collected in a baghouse or electrostatic precipitator (ESP).	Reduces the outlet flue gas SO_2 concentration to 75-100 ppmv.	The process has been tested on a demonstration size scale.	Same as for the lime dry scrubbing process.	Same as for the lime dry scrubbing process. Also, soda ash is an expensive raw material. Sulfur is nonrecoverable.
Oil Shale[b] (Processed Shale, Nahcolite)	Flue gas is contacted with an atomized absorbent slurry in a spray dryer scrubber. The alkaline minerals in the shale (primarily calcium and sodium carbonates) absorb the SO_2, are dried, and then collected in a baghouse or electrostatic precipitator (ESP).	Reduces the outlet flue gas SO_2 concentration to ~100-150 ppmv.	The process is only conceptual at this time and has not been tested on a pilot plant scale; however, the Lurgi oil shale retorting process lift pipe and flue gas treating equipment and the TOSCO II preheat unit closely resemble this system.	Same as for the lime dry scrubbing process. Also, an abundant supply of processed shale is available at the plant site. Nahcolite is plentiful in the Piceance Basin.	Same as for the lime dry scrubbing process. Sulfur is nonrecoverable.

* Performance somewhat depends on inlet SO_2 concentrations, fuel quantities and reagent utilization.

Sources: SWEC based on information from:

[a] Kohl and Riesenfeld, 1979.

[b] Stone and Webster Engineering Corp., January 30, 1979.

[c] Electric Power Research Institute, April 1980.

Wet scrubbing--The regenerable wet scrubbing processes generally employ a clean alkaline solution to absorb SO_2 in a scrubber. The resulting spent solution is treated with an insoluble alkali makeup which precipitates the absorbed SO_2. The insoluble alkali sulfite and sulfate crystals are then separated from the regenerated solution in a clarifier and possibly a second dewatering step such as a centrifuge. The spent alkali sludge is treated by calcining, evaporation, stripping, etc., which drive off the SO_2. The SO_2 can then be converted to a useful form of sulfur such as sulfuric acid or elemental sulfur.

In the nonregenerable processes, this spent alkali sludge is sent to a disposal area for land filling.

Dry scrubbing--The dry scrubbing processes use a concentrated slurry of alkaline crystals which are atomized and injected into the flue gas stream as it passes through a spray dryer. The scrubbing slurry absorbs the SO_2 and is dried by the hot flue gases. The dried spent alkali is then removed from the flue gas by an electrostatic precipitator or a baghouse.

In the regenerable processes, the spent alkali is reduced to a sulfide and then reacted with CO_2 to regenerate the alkali and evolve H_2S gas. The regenerated alkali is recycled, while the H_2S gas may be converted to elemental sulfur in a sulfur recovery unit.

In the nonregenerable processes, the spent material is sent to a disposal area for landfilling. The spent material also may be recycled to increase alkali utilization.

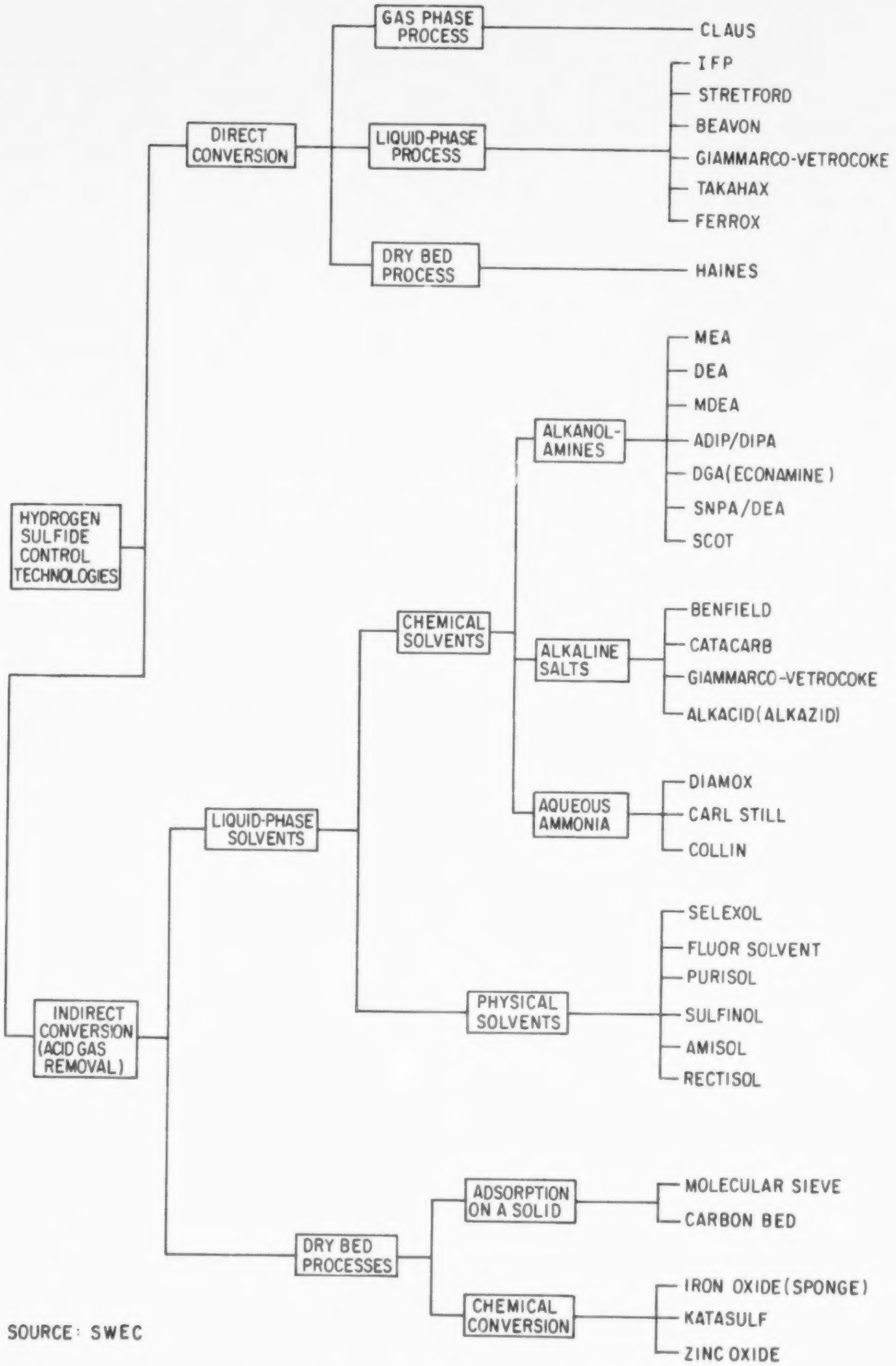
Hydrogen sulfide control. H_2S removal can be divided into two categories:

- Direct conversion
- Indirect conversion.

Direct conversion actually oxidizes H_2S to elemental S. Indirect conversion involves removing acid gases (H_2S and CO_2) from the gas stream and requires downstream direct conversion or further processing to treat the sulfur compounds. Figure 5.1-5 lists the H_2S removal systems available, and Table 5.1-9 presents a brief description of the process technologies.

Direct conversion--As shown in Table 5.1-9, several direct conversion technologies are currently available, including Claus, IFP, Stretford, Beavon, Giammarco-Vetrocoke, Takahax, Ferrox and Haines. The conversion of H_2S to elemental sulfur takes place in the liquid-phase in all the processes, except the Claus and Haines which are dry gas-phase removal processes.

Liquid-phase direct conversion processes are ideal for treatment of gases containing low concentrations of H_2S. In these processes, the acid gas components are absorbed by alkali solutions and then oxidized with dissolved oxygen to elemental sulfur. High circulation rates of the alkali solution are required for high performance and to reduce thiosulfate precipitate



HYDROGEN SULFIDE CONTROL TECHNOLOGIES
DIRECT CONVERSION
GAS PHASE PROCESS
CLAUS
LIQUID-PHASE PROCESS
IFP
STRETFORD
BEAVON
GIAMMARCO-VETROCOKE
TAKAHAX
FERROX
DRY BED PROCESS
HAINES
INDIRECT CONVERSION (ACID GAS REMOVAL)
LIQUID-PHASE SOLVENTS
CHEMICAL SOLVENTS
ALKANOL-AMINES
MEA
DEA
MDEA
ADIP/DIPA
DGA(ECONAMINE)
SNPA/DEA
SCOT
ALKALINE SALTS
BENFIELD
CATACARB
GIAMMARCO-VETROCOKE
ALKACID(ALKAZID)
AQUEOUS AMMONIA
DIAMOX
CARL STILL
COLLIN
PHYSICAL SOLVENTS
SELEXOL
FLUOR SOLVENT
PURISOL
SULFINOL
AMISOL
RECTISOL
DRY BED PROCESSES
ADSORPTION ON A SOLID
MOLECULAR SIEVE
CARBON BED
CHEMICAL CONVERSION
IRON OXIDE (SPONGE)
KATASULF
ZINC OXIDE

SOURCE: SWEC

FIGURE 5.1-5 HYDROGEN SULFIDE CONTROL TECHNOLOGIES

TABLE 5.1-9. KEY FEATURES OF HYDROGEN SULFIDE CONTROL TECHNOLOGIES

Control Technology	Operating Principle	Performance	Development Status	Advantages	Disadvantages
DIRECT CONVERSION					
Claus[a,b]	Partial oxidation of H_2S to SO_2 and subsequent reaction $2H_2S + SO_2 \rightarrow S + 2H_2O$ in gas-phase.	95% removal of H_2S.	Continuously improved designs available.	Provides extremely good quality elemental sulfur. Partially removes COS.	Side reactions with CO_2 and light hydrocarbons result in stable sulfur compounds emitted from the process. Tail gas may need to be treated.
IFP[b]	Absorption of H_2S and SO_2 (Claus tail gas) in polyalkene glycol, followed by conversion to elemental sulfur using a catalyst (liquid phase Claus reaction).	Reduces sulfur species to <1,500 ppmv.	Available for desulfurization of Claus tail gas.	Presence of high concentration of CO_2 has no adverse effect.	Low removal efficiency.
Stretford[a,b]	H_2S absorption and liquid-phase oxidation $H_2S + \frac{1}{2}O_2 \rightarrow S + H_2O$ in an alkaline solution of a vanadium salt.	Reduces the outlet H_2S concentration to 30 ppmv.	Process commercially available.	Process suitable for desulfurization of a variety of gas streams.	CO_2 absorbed in the process causes significant increases in absorber height requirements. HCN in feed produces a nonregenerable compound with high pollution potential.
Beavon[a,b]	Catalytic hydrogenation and hydrolysis of all sulfur compounds to H_2S, followed by recovery of elemental sulfur using the Stretford process (see above).	Reduces the outlet H_2S concentration to <50 ppmv.	Process commercially available.	Removes all sulfur compounds.	Same as the Stretford process (see above).
Giammarco-Vetrocoke[b]	H_2S absorption and liquid-phase oxidation $H_2S + \frac{1}{2}O_2 \rightarrow S + H_2O$ in a solution of arsenic salt.	99.99% removal of H_2S.	Available for desulfurization of coke-oven and synthesis gases and natural gas.	Capable of producing purified gas containing less than 1 ppmv H_2S even at temperatures up to 300°F.	Hazardous nature of arsenic solution.
Takahax[b]	H_2S absorption and liquid-phase oxidation $H_2S + \frac{1}{2}O_2 \rightarrow S + H_2O$ in an alkaline solution of naphthoquinone compounds.	99.99% removal of H_2S.	More than 100 units operating in Japan.	Capable of producing treated gas containing no detectable H_2S even at high inlet concentrations.	Sulfur precipitation is of very fine grain and amenable to removal via flotation techniques.

(Continued)

TABLE 5.1-9 (cont.)

Control Technology	Operating Principle	Performance	Development Status	Advantages	Disadvantages
Ferrox[b]	H_2S absorption and liquid-phase oxidation $H_2S + \frac{1}{2}O_2 \rightarrow S + H_2O$ in a solution of Na_2CO_3 and $Fe(OH)_3$.	85-99% removal of H_2S.	Few ferrox plants are still in operation.	Marked improvement over dry-box purification due to reduced installation and labor costs.	Sulfur from the ferrox process is not suitable for most uses and chemical replacement costs are high.
Haines[b]	Molecular sieves remove H_2S and water. H_2S is stripped from the bed and reacted with SO_2 to form elemental sulfur.	Reduces the outlet H_2S concentration to <5 ppmv.	Pilot plants in operation in Canada.	Selective adsorption of H_2S in the presence of CO_2. Also, removes mercaptans.	Zeolite adsorption beds may become fouled, impairing regeneration.
INDIRECT CONVERSION					
MEA[b]	H_2S and CO_2 absorbed by a regenerable reaction with Monoethanolamine at ambient temperatures.	Reduces the outlet H_2S concentration to 30 ppmv.	Used almost exclusively for years to remove H_2S and CO_2 from natural and certain synthesis gases.	Simultaneous removal of H_2S and CO_2. Applicable to low concentrations of H_2S and CO_2.	Nonselective for H_2S (e.g., CO_2 also absorbed). Reacts irreversibly with COS and CS_2.
DEA[b]	H_2S and CO_2 absorbed by a regenerable reaction with Diethanolamine at ambient temperature.	99% removal of H_2S.	Preferred choice for treatment of high-pressure natural gas with high concentrations of COS and CS_2.	Predominantly used in refining or manufacturing. Does not absorb COS and CS_2.	Nonselective for H_2S. Reclaiming requires vacuum distillation.
MDEA[a,b]	Selective absorption of H_2S by a regenerable reaction with Methyldiethanolamine.	99% removal of H_2S, 30-65% removal of CO_2.	Commercial plant under construction.	Higher selectivity to H_2S than primary or secondary amines.	Reactions between MDEA and HCN are irreversible.
ADIP/DIPA[a,b]	Selective absorption of H_2S by a regenerable reaction with Diisopropanolamine.	Reduces the outlet H_2S concentration to <100 ppmv.	More than 100 plants constructed worldwide.	Selective for H_2S removal. Substantial amounts of COS removed without detrimental effects. Low regeneration steam requirements.	Can require long residence times for sufficient removal.
DGA[b] (Econamine)	Absorption of H_2S by a regenerable reaction with Diglycolamine.	99% removal of H_2S.	Sour gas processing in operation.	DGA similar to DEA with lower vapor pressure. Lower circulation rates and steam consumption than DEA.	DGA costs are high. High corrosivity. Losses due to reaction with CO_2, COS and CS_2 are high. Reclaiming requires vacuum distillation.

(Continued)

TABLE 5.1-9 (cont.)

Control Technology	Operating Principle	Performance	Development Status	Advantages	Disadvantages
SNPA/DEA[b]	Process utilizes experience gained by SNPA in using DEA.	99% removal of H_2S.	Widely accepted choice for the treatment of high-pressure natural gases with high concentrations of acidic components.	COS and CS_2 are not harmful to the solution. Decomposition products removed by filtering through activated carbon.	High pressure process. Vacuum distillation probable.
SCOT[a,b]	Process uses DIPA to absorb H_2S from the Claus tail gas. Non-H_2S sulfur compounds are converted to H_2S before absorption in DIPA. Regenerated H_2S is oxidized to SO_2 before venting.	Reduces the outlet SO_2 concentration to 300 ppmv.	Several commercial units in operation since 1972.	Removes all sulfur compounds. Specifically suited for the Claus tail gas cleanup.	DIPA removes ~30% CO_2 which is recycled to the Claus process, diluting the feed.
Benfield[a,b] (Hot Potassium Carbonate)	Process uses hot (190°F) potassium carbonate to absorb CO_2 and H_2S. Regenerated by pressure reduction.	90-98% removal of H_2S, 10-40% removal of CO_2.	Process utilized worldwide, with further developments in recent years.	High temperature permits use of highly concentrated solution. Process is very simple. COS, CS_2 and RSH easily removed.	High pressure process.
Catacarb[b]	Similar to Benfield with the use of a proprietary catalytic additive to the hot $KHCO_3$.	Reduces the outlet H_2S concentration to <5 ppmv.	Plants in operation worldwide.	Higher gas purity and lower steam consumption than Benfield. Lower capital cost.	High pressure process.
Giammarco-Vetrocoke	SEE DIRECT CONVERSION PROCESSES				
Alkacid[a,b] (Alkazid)	Process uses various proprietary absorption solutions of alkaline salts and organic acid.	Reduces the outlet H_2S concentration to <5 ppmv.	Although operated abroad since the 1930's, no known commercial installations exist in the U.S.	Solutions are relatively noncorrosive. Solution tailored to requirements for H_2S selectivity and to minimize effect of contaminants.	May require special alloys to handle hot solutions. High CO_2 concentrations result in lower H_2S removal efficiency.
DIAMOX[b]	Selective H_2S removal process using absorption characteristics of ammonia with total (0.7 wt% NH_3) liquid recycle.	Reduces the outlet H_2S concentration to 100 ppmv.	Recently developed and commercialized in Japan.	Acid gas produced is suitable Claus feed or sulfuric acid plant feed. Low pressure process.	Purge stream of ammonia liquor is produced.
Carl Still[a,b]	Selective H_2S removal process using ammonia for absorption (2.0 wt% NH_3) with total liquid recycle.	Reduces the outlet H_2S concentration to ~560 ppmv.	Commercial process now in operation in the U.S.	Low pressure process. Good Claus plant feed.	H_2S selectivity less than DIAMOX. Concentrated NH_3 solutions are highly corrosive. Organic sulfides are not removed.

(Continued)

TABLE 5.1-9 (cont.)

Control Technology	Operating Principle	Performance	Development Status	Advantages	Disadvantages
Collin[b]	Selective H_2S removal process using NH_3 for absorption with total liquid recycle in a six-stage spray tower.	Reduces the outlet H_2S concentration to ~2,000 ppmv.	Commercial process now in use in Europe.	Low pressure process. Good Claus plant feed.	Does not remove organic sulfur compounds.
Selexol[a,b]	Uses an anhydrous organic solvent dimethylether of polyethylene glycol which physically dissolves acid gases and is stripped by reducing pressure without adding heat.	Reduces the outlet H_2S concentration to <1 ppmv.	Few plants in operation for natural gas treatment and for synthesis and coal-derived gas purification.	Nontoxic solvent. Partially removes COS and organic sulfur compounds.	Requires high partial pressure of acid gas.
Fluor Solvent[b]	Uses an anhydrous organic solvent proprylene carbonate which physically dissolves acid gases.	Reduces the outlet H_2S concentration to <5 ppmv.	Plants in operation for CO_2 gases and combination CO_2 and H_2S gases.	Low operation costs.	Solvent intended primarily for removal of CO_2.
Purisol[a,b]	Uses an anhydrous organic solvent N-methyl-2-pyrolidine which physically dissolves acid gases.	Reduces the outlet H_2S concentration to <4 ppmv and CO_2 concentration to 2-3 vol. %.	Four commercial installations in operation as of 1979.	High H_2S selectivity. Partially removes COS and organic sulfur compounds.	Operating temperatures must be near ambient and a certain minimum acid gas partial pressure is required.
Sulfinol[a,b]	Uses a mixture of chemical (DIPA) and physical solvent (sulfolane) and water.	Reduces the outlet H_2S concentration to <1 ppmv and CO_2 to <50 ppmv.	Wide application in the treatment of natural, refinery and synthesis gases.	Removes COS and RSH. Capacity of the sulfinol high at high partial pressure of H_2S.	Optimum operation requires high pressure. Solution absorbs heavy HC's, requiring flash tank separation.
Amisol[b]	Uses a mixture of a chemical (MEA/DEA) and a physical solvent (methanol).	Reduces the outlet H_2S concentration to <0.1 ppmv and CO_2 to <5 ppmv.	Only semi-commercial plants in operation.	Capable of removing all sulfur compounds.	Nonselective for H_2S.
Rectisol[a,b]	Uses physical absorption in methanol at low temperature.	Reduces the outlet H_2S concentration to <0.1 ppmv.	Thirty-six plants are in operation worldwide. Twelve units are under construction.	Heat input low because temperature is maintained by flashing. Removes all undesirable components (COS, CS_2, RSH, HCN) in a single step.	Complex operation. High solvent losses. Best suited for higher pressures. Low temperature process.
Molecular Sieve[a,b]	Use of molecular sieves to adsorb sulfur compounds.	Reduces the outlet H_2S concentration to <4 ppmv.	Not widely used for removing H_2S from gas streams.	Extended useful life (3-5 years) of adsorbent possible with properly designed molecular sieve. Removes mercaptans.	Preferably used on high pressure streams. Regeneration gas disposal.

(Continued)

TABLE 5.1-9 (cont.)

Control Technology	Operating Principle	Performance	Development Status	Advantages	Disadvantages
Carbon Bed[a,b]	Activated carbon catalytically oxidizes H_2S to elemental sulfur at ambient temperatures. Sulfur removed by solvent washing.	Reduces the outlet H_2S concentration to 0.2 ppmv.	Sixty commercial plants in operation in the U.S.	Very pure sulfur is obtained. Complete H_2S removal.	Carbon deactivated rapidly. Purification required to remove tar and ammonia. Complicated sulfur extraction procedures. 1,500 ppmv/H_2S limit in feed.
Iron Oxide[a,b] (Iron Sponge)	H_2S removed completely by reaction with ferric oxide $2Fe_2O_3 + 6H_2S \rightarrow 2Fe_2S_3 + 6H_2O$. Exposure to air oxidizes Fe_2S_3 to Fe_2O_3 and sulfur.	Reduces the outlet H_2S concentration to 0.3 ppmv.	Very old process, still used on a large scale for the treatment of coal gases.	Low pressure drop process. Complete removal of H_2S.	Suitable for low volume/low pressure gases. Fouled Fe_2O_3 disposal is required. Tar and HCN contaminated Fe_2O_3.
Katasulf[b]	Air and preheated gas with H_2S catalyzed to form SO_2 which is absorbed in an aqueous ammonium sulfite-bisulfite solution.	Reduces the outlet H_2S concentration to <4 ppmv.	Large commercial units in operation worldwide.	Produces a salable ammonium sulfate.	Carbon-steel corrosion problems exist with some forms of the process.
Zinc Oxide[a]	The gas is passed through a bed of zinc oxide, resulting in the reaction $ZnO + H_2S \rightarrow ZnS + H_2O$.	Reduces the outlet H_2S concentration to 0.2-0.5 ppmv.	One hundred commercial plants in operation worldwide.	Virtually complete removal of H_2S.	ZnO is not regenerable and is expensive.

Sources: SWEC based on information from:

[a] Dravo Corp., February 1976.

[b] Kohl and Riesenfeld, 1979.

formation. High selectivity for H_2S removal can also be achieved by taking advantage of the higher H_2S versus CO_2 absorption rates.

The gas-phase direct conversion (Claus and Haines processes) consists of thermal oxidation of one-third of the H_2S to SO_2, followed by a series of catalytic reactors that react SO_2 with the remaining H_2S to form elemental sulfur. The heat for combustion in the furnace is obtained from the oxidation of H_2S; thus, the H_2S concentration must be high enough to sustain spontaneous combustion. Therefore, the gas-phase conversion requires an acid gas stream with a higher H_2S concentration than the liquid-phase conversion.

Indirect conversion--There are essentially five classes of commercially available, indirect H_2S removal technologies that are used in conjunction with direct conversion technologies; these are, removal of H_2S by:

I. Alkanolamine
II. Alkaline salts
III. Aqueous ammonia
IV. Physical solvents
V. Dry bed processes.

The alkanolamine processes (I) remove acidic impurities, i.e., H_2S, CO_2, COS, and CS_2, from gases by an acid-base chemical reaction with the amine. The process involves an absorption-regeneration cycle of a circulating amine. Commonly used amines are monoethanolamine (MEA), methyldiethanolamine (MDEA), diethanolamine (DEA), diisopropanolamine (DIPA) and diglycolamine (DGA). Major equipment systems used in the amine process are a gas-amine contactor (absorber) for absorption of the acid gases and a regenerator (stripper) for releasing the acid gas from solution. A downstream sulfur recovery facility is required to oxidize, or recover, the H_2S.

Alkaline salt processes (II) use an aqueous solution of a buffered potassium salt. The weak alkaline solution absorbs the acid gas components of the feed gas. The process operates at medium to high pressures because the absorption capability is influenced by the acid gas (H_2S and CO_2) partial pressures. The alkaline solution is regenerated by reducing the rich solution pressure to near ambient pressure, followed by steam stripping and sulfur recovery.

The ammonia process (III) uses the same mechanism for H_2S removal as the alkaline salt process (II) except the ammonia is used as the absorption agent. Regeneration and additional sulfur recovery are necessary.

Physical solvents (IV) have low heats of solution and can absorb acid gases in proportion to their partial pressures. These processes require high acid gas partial pressures which are achieved at low gas pressures and high acid gas concentrations, or at high gas pressures and low acid gas concentrations. Physical solvent processes are most economical when the feed gas is at high pressure and bulk removal of the acid components is desired. A high degree of selectivity of H_2S absorption is possible, but additional equipment is required, increasing costs. A downstream sulfur facility is also necessary to recover the H_2S.

Dry bed processes (V) generally employ two techniques to remove H_2S from a gas stream: (1) adsorption onto a dry bed, such as a molecular sieve or activated carbon, followed by desorption of the H_2S from the bed using a hot gas stream; and (2) reacting the H_2S with a dry bed material such as iron oxide to form a solid sulfide compound, which is then oxidized by air to regenerate the dry bed and to form elemental sulfur.

Sulfur Control Technologies Analyzed--

The fuel-based sulfur in the processed shale is the prime source of the SO_2 emissions from the Lurgi-Open Pit plant. The residual organic matter remaining on the oil shale after retorting is incinerated in the lift pipes, which results in the formation of SO_2. In addition, the hydrogen sulfide from the retort gas is removed so that the gas can be sold. The separated H_2S would also be a source of sulfurous emissions from the plant. Since the plant does not consume any fuels (except for the diesel fuel), there would be no additional SO_2 emissions.

The concentration of SO_2 in the flue gas from a 4,400 TPSD Lurgi module is estimated to be about 30 ppmv (Rio Blanco Oil Shale Co., February 1981). A comparable value of 20 ppmv has been estimated for the Lurgi commercial plant processing approximately 62,000 TPSD of oil shale at Tract C-b (Occidental Oil Shale, Inc. and Tenneco Shale Oil Co., April 1981). These values are very low for an indirectly retorted shale such as the Lurgi processed shale. As an explanation for the low value, Rio Blanco suggests that approximately 93% of the SO_2 formed during the processed shale incineration reacts irreversibly with the calcined material in the processed shale and the excess oxygen, forming stable sulfates. The possibility of this type of reaction has also been mentioned for the plant at Tract C-b and by Colony Development Operation (see the MIS-Lurgi and TOSCO II PCTMs, respectively). Since the SO_2 concentration in the Lurgi flue gas is reported to be at such a low level, SO_2 control technologies were not examined for the flue gas.

As stated earlier, instead of burning the retort gas in the plant, it is cleaned for selling purposes; therefore, its treatment by the DEA process is viewed as a processing, rather than pollution control, activity. However, the acid gases (H_2S, CO_2) from the DEA process, if emitted as such, would create pollution because approximately 700 lb/hr of H_2S (15.7 TPSD SO_2 equivalent) are contained in the gases. The Holmes-Stretford process was examined for the removal and recovery of H_2S from these acid gases. This process is selective in removing H_2S in the presence of CO_2. The acid gases have a CO_2 to H_2S ratio of 80:1 and only 6% of the CO_2 (approximately 3,300 lb/hr) is estimated to be absorbed by the Stretford solution; also, the H_2S concentration in the treated gas can be reduced to 30 ppmv, or 1.3 lb/hr (Peabody Process Systems, Inc., February 1981). The H_2S conversion reactions require the presence of large amounts of the Stretford chemicals, and absorption of CO_2 further increases the demand for these chemicals. Thus, at high absolute concentrations of both H_2S and CO_2, the Stretford process becomes less attractive due to the increased cost of solution circulation and regeneration.

Although the Stretford process can remove H_2S efficiently, this is not the case for other sulfur compounds such as COS, CS_2, mercaptans, etc. In general, all of these compounds have been found in oil shale retort gases, and Lurgi has indicated that COS is present in the Lurgi retort gas (Private communication with Hans Weiss, Lurgi Kohle und Mineralötechnik GmbH, West Germany, January 1981).

Table 5.1-10 gives equipment and cost details for the Holmes-Stretford unit. Since this is the only point of sulfur control in the Lurgi plant, the table also includes the cost of sulfur control. A specific cost curve based on the design of the Stretford unit used in this manual is presented in Figure 5.1-6. The description and stream characteristics for the Stretford process can be found in Sections 3 and 4.

TABLE 5.1-10. MAJOR ITEMS IN THE HOLMES-STRETFORD PROCESS[a]

Capital Cost Items	Operating Cost Items
Knock-Out Drum 3' diameter × 7'	Holmes-Stretford Mix 16 lb/day
Absorber 3' diameter × 65'	Soda Ash 607 lb/day
Oxidizers (6) 15' diameter × 19'	Process Water 3 gpm
Pump Tank 26' diameter × 14'	Steam 207 lb/hr
Circulation Pumps (3) 1,700 gpm @ 120 HP	Cooling Water 14 gpm
Flash Drum 3' diameter × 7'	Electricity 600 kW
Slurry Tank 20' diameter × 40'	Manpower 3 Men/day
Slurry Pumps (2) 75 gpm	Direct Annual Operating Cost, $\$10^3$ Maintenance 134 Operating Supplies 164
Filter System 250 lb/hr	Labor 350 Utilities 121 TOTAL 769
Sulfur Melter 200,000 Btu/hr	

(Continued)

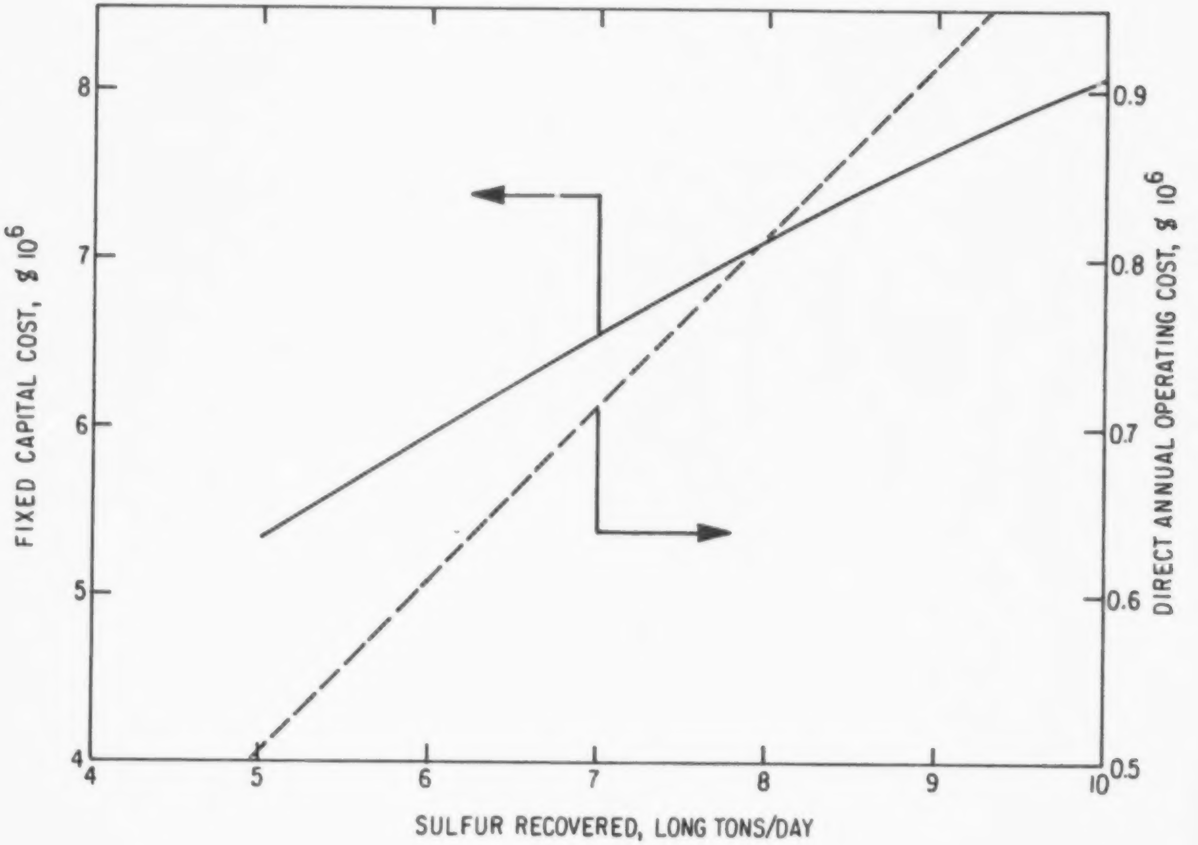
TABLE 5.1-10 (cont.)

Capital Cost Items	Operating Cost Items
Sulfur Decanters (2) 14' diameter × 14'	
Sulfur Storage Pit 75-ton capacity	
Evaporator 250 gpm liquor feed	
Heater 6,000 Btu/hr	
Cooler 4,000 Btu/hr	
Feed Gas Booster 90,000 ACFM @ 0 psig	
Flash Gas Boosters (2) 2,000 ACFM	
Plot Area 87,000 ft^2	
Site Preparation and Foundations	
Ductwork and Piping	
Electrical	
Instrumentation and Controls	
Painting	
Fixed Capital Cost, $\$10^3$ 6,860	
Total Annual Operating Cost, $\$10^3$ 915	
Total Annual Control Cost,[b] $\$10^3$ (¢/bbl) 2,044 (9.9)	

[a] Design basis: 10,500 ACFM, 7.6 LTPSD sulfur recovered.

[b] See Section 6 for details on computation of the total annual control cost.

Source· SWEC estimates based on information from Peabody Process Systems, Inc., February 1981.



SOURCE: DRI based on information provided by SWEC

FIGURE 5.1-6 COST OF SULFUR RECOVERY WITH STRETFORD PROCESS

Total Sulfur Emissions--

Sulfur dioxide, as such, is emitted only from the Lurgi flue gas discharge system and diesel equipment. In addition, the Stretford tail gas emits H_2S. These three sources constitute the total sulfur emissions from the plant and these are listed in Table 5.1-11.

TABLE 5.1-11. TOTAL SO_2 EMISSIONS FROM THE PLANT

Stream Number	Emission Source	Control Description	SO_2 Emissions (lb/hr)
24	Diesel Equipment	--	35.6
31	Flue Gas Discharge	--	500.9[a]
63	DEA Unit	Stretford	2.4[b]
TOTAL			538.9

[a] According to the information from Rio Blanco Oil Shale Co., February 1981, the control of SO_2 occurs in the lift pipes by adsorption on the processed shale. Approximately 93% of the SO_2 is claimed to be adsorbed, resulting in an SO_2 concentration of 30 ppmv in the flue gas.

[b] According to the information from Peabody Process Systems, Inc., February 1981, H_2S in the acid gases from DEA is reduced to a level of 30 ppmv. The value given above is the SO_2 equivalent from 1.3 lb/hr of H_2S emitted in the treated gases.

Source: SWEC estimates, except as noted.

5.1.3 Nitrogen Oxides Control

In oil shale processes, nitrogen enters the system from two primary sources: (1) the fuels derived from the raw shale, and (2) the air required for combustion in the various furnaces, heaters, auxiliary boilers and incinerators. A portion of this nitrogen is converted into other forms such as nitrogen oxides (NOx) and ammonia (NH_3). The NOx produced during fossil fuel combustion are emitted as NO and NO_2 in flue gases. These compounds are formed from the oxidation of nitrogen compounds (e.g., ammonia, cyanides) in the shale-derived fuels and/or from the fixation of atmospheric nitrogen (N_2). A large portion of ammonia resulting from the pyrolysis of the shale is usually removed in the gas condensate, or foul water, when the retort gas is cooled or scrubbed with water. This removal and subsequent recovery of ammonia provide an indirect control over NOx emissions. Since the recovery of ammonia from an aqueous solution also constitutes water pollution control, this aspect of the NOx control is discussed under water management (Section 5.2). The portions of ammonia and fuel-based nitrogen that are not removed in the gas condensate may require removal or control prior to emission to the environment. Federal and Colorado State standards and regulations limit NOx emissions because of their potential role in the formation of photochemical smog and acid precipitation.

Inventory of Control Technologies--

There are three categories of NOx control technologies:

- Reduction of nitrogen in the fuel
- Combustion modifications
- Stack gas removal.

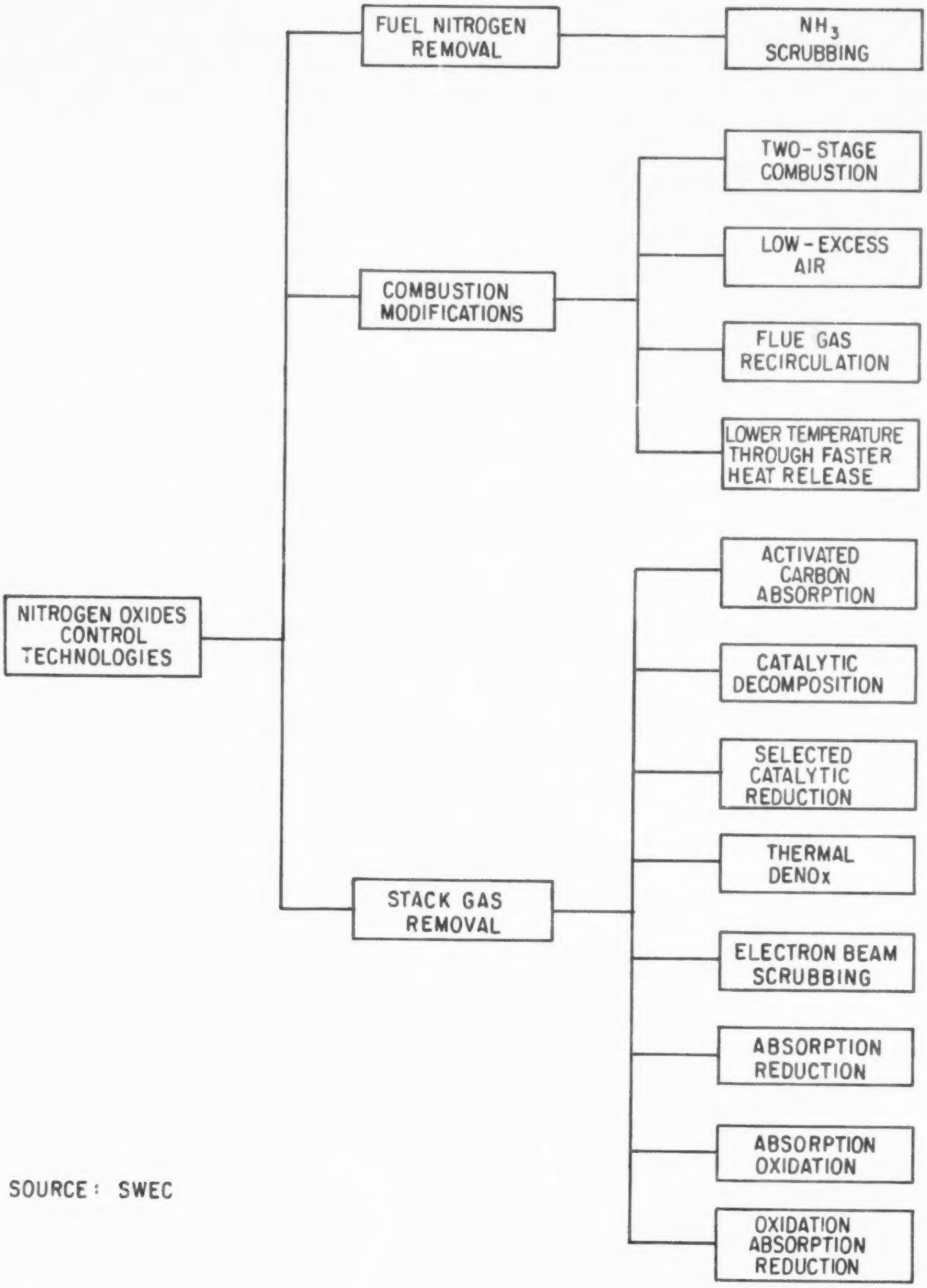
These processes are shown in Figure 5.1-7 and are discussed briefly in Table 5.1-12.

Reduction of nitrogen in the fuel. Burning fuels low in nitrogen is the simplest method of controlling NOx emissions arising from fuel-based nitrogen. Hydrotreatment of fuel oils and water scrubbing of fuel gases are fairly effective in removing the fuel-based nitrogen.

Combustion modifications. The generation of NOx by thermal fixation of atmospheric nitrogen is dependent upon the flame temperature, concentration of nitrogen, time history of individual combustion gas pockets, and the amount of excess air present. To some extent, these variables are controllable, and the production of NOx can be minimized for a particular combustion process.

Combustion control of NOx may be accomplished by several methods. One approach is design and operation of burners with fuel-rich mixture ratios. This technique, called off-stoichiometric combustion, produces low flame temperatures and, hence, potentially low NOx formation. A significant excess of oxygen is avoided in the combustion zone by diverting some portion of the inlet air through remote locations in the burner or through entirely separate secondary combustion air ports.

Another NOx reduction technique, based on combustion modification, takes advantage of the strong temperature dependency of nitric oxide (NO) formation on peak combustion temperatures. Reduced flame temperatures may be obtained by direct reduction of gas temperature or by indirectly increasing heat transfer. Direct techniques include recirculating product flue gases back into the combustion zone where they serve as diluents absorbing heat, thereby reducing maximum flame temperatures achieved. Other direct techniques are reduced combustion air preheat and water/steam injection. The latter is more applicable to gas turbines. Indirect NOx reduction relating to the combustion process usually involves furnace designs with increased burner spacing and heat removal capability. Flame temperature reduction does not reduce NOx formation from fuel nitrogen but does reduce atmospheric N_2 fixation.


NITROGEN OXIDES CONTROL TECHNOLOGIES
FUEL NITROGEN REMOVAL
NH_3 SCRUBBING
COMBUSTION MODIFICATIONS
TWO-STAGE COMBUSTION
LOW-EXCESS AIR
FLUE GAS RECIRCULATION
LOWER TEMPERATURE THROUGH FASTER HEAT RELEASE
STACK GAS REMOVAL
ACTIVATED CARBON ABSORPTION
CATALYTIC DECOMPOSITION
SELECTED CATALYTIC REDUCTION
THERMAL DENOx
ELECTRON BEAM SCRUBBING
ABSORPTION REDUCTION
ABSORPTION OXIDATION
OXIDATION ABSORPTION REDUCTION

SOURCE: SWEC

FIGURE 5.1-7 NITROGEN OXIDES CONTROL TECHNOLOGIES

TABLE 5.1-12. KEY FEATURES OF NITROGEN OXIDES CONTROL TECHNOLOGIES

Control Technology	Operating Principle	Performance	Development Status	Advantages	Disadvantages
FUEL NITROGEN REMOVAL					
NH_3 Scrubbing	Absorption of NH_3 by countercurrent scrubbing with water.	Up to 100% of NH_3 removal possible by changing water rate, composition and temperature.	Commercially proven.	Removes source of NOx before formed. By-product NH_3 is produced.	Does not reduce NOx emissions formed by thermal fixation of nitrogen and oxygen in combustion air.
COMBUSTION MODIFICATIONS					
Two-Stage Combustion (either low-emission burners or engineered combustion box)	Air is introduced in two zones. Zone 1) combustion occurs under reducing conditions; Zone 2) additional air added to complete combustion.	40-60% of thermal NOx reduction. Less reduction for fuel-nitrogen.	Burners and boiler designs commercially available.	Burner system is inexpensive in relation to total cost.	Reducing zone can cause boiler tube damage.
Low-Excess Air	Reduce excess air available to reduce reaction kinetics of N-radical and O_2 reaction.	10-20% NOx reduction.	Commercially proven.	Requires only operational changes.	Boiler is more difficult to operate. Possible increase in CO/HC emissions.
Flue Gas Recirculation (FGR)	Some flue gas recirculated to combustion zone to reduce peak temperature in furnace.	Up to 20% NOx reduction.	Commercially proven.	Useful in controlling outlet steam temperatures.	High capital and operating cost.
Lower Temperature Through Faster Heat Release	A larger boiler is used to reduce temperature in the furnace.	Up to 20% NOx reduction.	Commercially proven.	Results in significant reduction of NOx.	Difficult to implement in existing plants.
STACK GAS REMOVAL OF NOx					
Activated Carbon Adsorption	Adsorption of NOx (and SO_2) on activated carbon.	70-100% NOx, 72-90% SO_2 removal.	Sumitomo Huy Ind -1- plant discontinued operation due to high cost.	Simultaneous removal of SO_2 and NOx.	High consumption of carbon and fuel.
Catalytic Decomposition	No data available.		R.M. Parsons conceptual stage.		
Selective Catalytic Reduction (SCR)	NOx reduced by NH_3 over a catalyst (all processes similar using various proprietary catalysts).	Up to 90% NOx removal.	Four processes in commercial-scale operation, 20 processes available.	Low operating temperatures and low consumption of NH_3.	Particulates and SO_3 can cause catalyst plugging and poisoning. High capital cost.

(Continued)

TABLE 5.1-12 (cont.)

Control Technology	Operating Principle	Performance	Development Status	Advantages	Disadvantages
Thermal DeNOx	NH_3 injected in a 1,300-1,800°F flame zone where $NO + NH_3 \rightarrow N_2 + H_2O$.	Up to 70% NOx removal.	Demonstrated commercially.	By-product recovery not required. Low capital cost.	Requires large amounts of NH_3. Narrow operating range.
Electron Beam Scrubbing	Removal of SO_2 and NOx by reaction with NH_3 in the presence of electrons. Products are $(NH_4)_2SO_4$ + NH_4NO_3.	Up to 85% NOx, 90% SO_2 removal.	Pilot plant stage only.	Simultaneous removal of SO_2 and NOx.	Power consumption is 3% of plant output for beam accelerator. High capital cost. Requires high efficiency ESP.
Absorption Reduction	NOx is converted to NH_3 by the reducing effect of SO_2 to make $(NH_4)_2SO_4$ with a liquid Fe EDTA catalyst in a 20-tray column.	70-85% NOx, 90% SO_2 removal.	Not demonstrated commercially.	No oxidizing agent required.	Requires an expensive column. Consumes large quantities of NH_3 and H_2SO_4.
Absorption Oxidation	NOx and SO_2 are absorbed in a $KOH/KMnO_4$ solution and are oxidized to KNO_3 and K_2SO_4.	No data available.	Not demonstrated commercially.	Simultaneous removal of SO_2 and NOx.	Costly due to consumption of KOH and regeneration of $KMnO_4$.
Oxidation Absorption Reduction	Either O_3 or ClO_2 are used to oxidize NO to NO_2.	Up to 85% NOx, 95% SO_2 removal if SO_2/NOx ratio is 2.5.	Not demonstrated commercially.	Simultaneous removal of SO_2 and NOx.	Chloride in the waste stream causes disposal problems. Nitrate formation can also cause disposal problems. Oxidation material is very expensive.

Source: SWEC based on information from Battelle, Columbus Laboratories, October 1978.

Stack gas removal. Flue gas treatment for NOx removal is a relatively new, developing technology. Two broad categories may be defined: wet processes in which NOx is absorbed into an aqueous solution, and dry processes in which NOx is reduced by ammonia.

The wet NOx removal processes also serve as a mechanism to reduce sulfur dioxide emissions and, as such, can provide effective environmental control where both pollutants are present. However, due to the low solubility of NOx in aqueous solutions and the low removal efficiencies obtainable, absorption techniques usually prove to be very expensive.

Dry NOx removal systems, in general, display higher nitrogen oxide reduction capabilities and are economically more viable than wet systems. These processes are usually ammonia based and may be selective or nonselective and catalytic or noncatalytic. Depending on the individual process applied, ammonia is injected into the flue gas at some point after complete combustion and prior to a minimum gas temperature of 350°F. In the resulting reaction, NOx combine with ammonia to form molecular nitrogen and water.

Nitrogen Oxides Control Technologies Analyzed--

The primary source of NOx emissions from the Lurgi-Open Pit plant is the Lurgi flue gas discharge system. According to Rio Blanco, the NOx emissions in the Lurgi flue gas originate only from the fuel-based nitrogen in the organic residue on the processed shale. The temperature in the lift pipes (1,240°F) is claimed to be low enough so that thermal fixation of the atmospheric nitrogen does not occur during processed shale incineration (Rio Blanco Oil Shale Co., February 1981). Since there is no fuel combustion in the plant, additional NOx emissions are not formed.

Ammonia in the Lurgi retort gas is removed during product liquor condensation. Since this is an integral processing step in the Lurgi technology, it is not considered a pollution control measure.

Once removed in the Lurgi gas liquor, the actual recovery of NH_3 is achieved with an ammonia recovery process. Since the process is considered to be a water treatment technology, it is discussed in Section 5.2.

The modified DDP for Tract C-a (Rio Blanco Oil Shale Co., February 1981) reports that the concentration of NOx in the flue gas is 300 ppmv, which is equivalent to 3,600 lb/hr of NO_2, or 2,430 lb/hr NOx assuming 90% NO and 10% NO_2, by weight. In a separate organic nitrogen material balance presented in the same document, however, 0.3 lb of organic nitrogen/ton of raw shale is reported to be converted to NOx. This latter value is equivalent to about 4,900 lb/hr NO_2, or 400 ppmv in the flue gas.

If the formation of NOx in the flue gas is due only to the oxidation of fuel-based nitrogen, as is claimed by Rio Blanco (Rio Blanco Oil Shale Co., February 1981), combustion modifications cannot be employed to control the NOx. Also, techniques do not exist for removing organic nitrogen from the processed shale. However, if thermal fixation of atmospheric nitrogen does occur in the lift pipes, combustion modifications can be applied in order to reduce the NOx formation. The stack gas NOx removal techniques may be applicable regardless of the origin of NOx, but most of them have not been successful in commercial-scale, continuous operations. Only refrigerated tanks for the storage of ammonia were examined as an indirect NOx control measure. The fixed capital cost for the storage tanks is estimated to be $466,000 and the total annual operating cost is $15,000. This results in a total annual control cost of $88,000, or 0.4 cents/bbl of oil (see Section 6 for details on computation of the total annual control cost). The cost for the ammonia storage tanks also constitutes the total cost of NOx control for the plant.

Total Nitrogen Oxides Emissions--

There are only two plant emissions that contain NOx--the flue gas and diesel exhaust. The quantities of NOx in the two streams are listed in Table 5.1-13.

TABLE 5.1-13. TOTAL NOx EMISSIONS FROM THE PLANT

Stream Number	Emission Source	NOx Emissions[a] (lb/hr)
24	Diesel Equipment	469.9
31	Flue Gas Discharge System	2,432.4[b]
TOTAL		2,902.3

[a] Expressed as 90% NO and 10% NO_2, by weight.

[b] Value is based on 300 ppmv NOx in the flue gas, according to the information from Rio Blanco Oil Shale Co., February 1981.

Source: SWEC estimates, except as noted.

5.1.4 Hydrocarbon Control

Hydrocarbon compounds are emitted to the atmosphere as a result of incomplete fuel combustion or as a fugitive emission from small leaks in processing or storage equipment.

The hydrocarbon emissions from noncombustion sources are usually referred to as volatile organic compounds (VOC) or reactive hydrocarbons (RHC) in government regulations restricting their emission. Federal and State regulations limit these hydrocarbon emissions because of their role in the formation of photochemical smog and ozone production.

Inventory of Control Technologies--

As illustrated in Figure 5.1-8 and discussed in Table 5.1-14, hydrocarbon emissions can be controlled by the following categories of control technologies:

- Additional sealing of process equipment
- Vapor recovery
- Complete fuel combustion
- Catalytic converters
- Thermal oxidizers.

Additional sealing of process equipment. Hydrocarbon emission control by additional sealing of process and storage equipment is best accomplished by engineering these features into the plant. This includes double seals on tanks, pumps, and other rotating machinery, closed-loop sampling, caps on open-ended valves, and periodic monitoring of equipment to find hydrocarbon leaks quickly. This will result in a minimum additional plant capital cost and will more than pay for itself due to the value of the hydrocarbons which are prevented from being emitted.

Vapor recovery. When hydrocarbon vapor emissions cannot be controlled by additional sealing of equipment, a vapor recovery system can be installed to collect and condense the vapors by refrigeration and return them to the process.

Complete fuel combustion. The most cost-effective way to control hydrocarbon emissions from fuel combustion processes is to operate the process with enough excess air to ensure complete oxidation of all hydrocarbons to CO_2 and H_2O, i.e., complete fuel combustion.

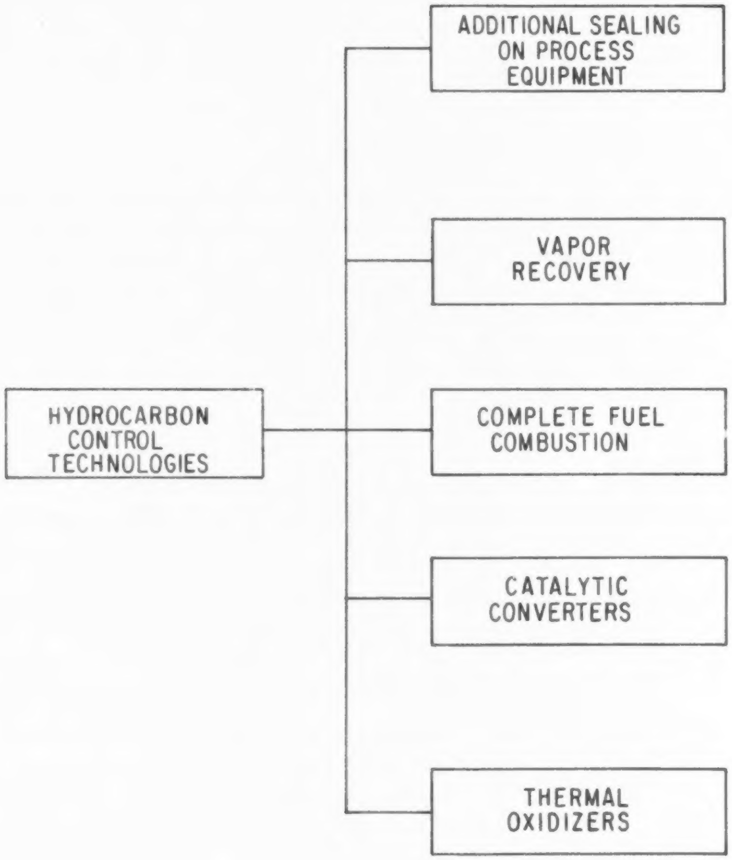
Catalytic converters. When complete fuel combustion does not occur, the hot exhaust gas from the process can be sent through a catalytic converter. In the catalytic converter, the gas is passed over a catalyst where the unburned hydrocarbons are reacted with the excess air in the exhaust gas and are converted to CO_2 and H_2O.

Thermal oxidizers. Hydrocarbon vapor streams or any other waste gas stream containing unburned hydrocarbons can be burned in a thermal oxidizer with excess air and additional fuel, if needed; this completely oxidizes all hydrocarbons to CO_2 and H_2O.

Hydrocarbon Control Technologies Analyzed--

The hydrocarbon emissions in the Lurgi-Open Pit plant emanate from the leakage in the valves, pumps, etc., as the fugitive emissions from oil product storage, and due to the incomplete combustion of the fuels.

Hydrocarbon emissions from diesel-burning equipment are controlled by installation of catalytic conversion systems. The least costly fugitive


HYDROCARBON CONTROL TECHNOLOGIES
ADDITIONAL SEALING ON PROCESS EQUIPMENT
VAPOR RECOVERY
COMPLETE FUEL COMBUSTION
CATALYTIC CONVERTERS
THERMAL OXIDIZERS

SOURCE: SWEC

FIGURE 5.1-8 HYDROCARBON CONTROL TECHNOLOGIES

TABLE 5.1-14. KEY FEATURES OF HYDROCARBON CONTROL TECHNOLOGIES

Control Technology	Operating Principle	Performance	Development Status	Advantages	Disadvantages
Additional Sealing on Process Equipment	Double seals on pumps and rotating machinery and caps on open-ended valves reduce hydrocarbon losses from the equipment.	About 60%-65% reduction of fugitive hydrocarbon emissions is possible with this level of control.	Commercially proven.	Requires a small capital and operating cost and will probably more than pay for this cost due to the value of the hydrocarbons which are prevented from being emitted.	Should be implemented during new plant construction. Requires more capital investment to retrofit the controls of an existing plant.
Vapor Recovery	Hydrocarbon vapors emitted from process equipment are collected and condensed by refrigeration and then returned to the process.	About 80-90% of the hydrocarbon vapors can usually be condensed and returned to the system.	Commercially proven.	A reliable system which is best applied to potential point source emission streams.	Can be a high energy requirement to operate the refrigeration system.
Complete Fuel Combustion	Combustion process is operated with excess air to ensure complete oxidation of all hydrocarbons to CO_2 and H_2O.	Can convert close to 100% of all hydrocarbons in the fuel to CO_2 and H_2O.	Commercially proven.	Eliminates the need for downstream equipment to complete the conversion of CO to CO_2.	An adequate air:fuel ratio must be maintained.
Catalytic Converters	Hot exhaust gas is passed over a catalyst where the unburned hydrocarbons are reacted with the excess air in the exhaust gas and are converted to CO_2 and H_2O.	Can convert up to 80% of the hydrocarbons in diesel exhaust gas streams to CO_2 and H_2O; for other fuel burning processes, up to 99% conversion is possible.	Commercially proven.	Does not require any fuel and has no moving parts so that routine maintenance is minimal.	The catalyst, which is expensive, must be replaced periodically.
Thermal Oxidizers	Waste gas streams containing unburned hydrocarbons are burned with excess air and additional fuel if needed to completely oxidize all hydrocarbons to CO_2 and H_2O.	Can convert close to 100% of all hydrocarbons in the gas stream to CO_2 and H_2O.	Commercially proven.	Will ensure complete oxidation of hydrocarbons and any other unwanted components in the gas stream.	Can have a high energy requirement when supplemental fuel is used.

Source: SWEC based on information from Research and Education Association, 1980.

hydrocarbon emissions control for storage tanks is proper sealing. Alternatively, vapor recovery can be used, but the expense is extremely high for these systems. As a standard industry practice, double-sealed, floating roof storage tanks are provided for volatile product storage. Internal plant leaks are controlled by use of adequate seals and strict maintenance procedures. Approximately 232 lb/hr of hydrocarbons (expressed as methane) are estimated by Rio Blanco for the 4,400 TPSD Lurgi module (Rio Blanco Oil Shale Co., February 1981). Except for using proper combustion practices, no other technologies are provided to reduce the hydrocarbon release in the flue gas.

Table 5.1-15 lists the hydrocarbon control practices and equipment considered, and Table 5.1-16 presents the costs for hydrocarbon control for the entire plant.

TABLE 5.1-15. HYDROCARBON CONTROL PRACTICES AND EQUIPMENT

Capital Cost Items	Operating Cost Items
Floating Roof Storage Tanks (2) 200' diameter × 48', 268,000 bbl (each) Welded API 650 code Double seals Carbon steel	Maintenance
Complete Combustion of Fuels	
Dual Mechanical Seals on Pumps and Valves	
Catalytic Converters on all Diesel Equipment	
Monitoring Equipment	

Source: SWEC.

Total Hydrocarbon Emissions--

Table 5.1-17 summarizes the hydrocarbon emission sources and control equipment used for the emissions.

5.1.5 Carbon Monoxide Control

Carbon monoxide (CO) is usually formed by incomplete combustion of fossil fuels. Normally, an excess of oxygen is supplied to a combustion process to ensure that all of the carbon in the fuel is converted to carbon dioxide (CO_2). When a shortage of oxygen occurs in the combustion process, some of the carbon is only partially oxidized to CO. Federal and State standards and regulations limit CO emissions because of their deleterious effect on the human respiratory system.

TABLE 5.1-16. COST OF HYDROCARBON POLLUTION CONTROL

Stream Number	Control Description	Control Location	Number of Units	Fixed Capital Cost ($000's)	Total Annual Operating Cost ($000's)	Total Annual Control Cost ($000's)[a]
24	Catalytic Converters	Diesel Equipment	--	170	65	106
112	Maintenance	Valves, Pumps, etc.	--	61	(59)[b]	(44)
44, 47	Floating Roof Storage Tanks	Product Storage	2	300	(141)	(89)
TOTAL				531	(135)	(27)

[a] See Section 6 for details on computation of the total annual control cost.

[b] Values in parentheses ( ) indicate profit after subtracting the total annual capital and operating charges from the annual by-product credit of $125,000 from maintenance and $155,000 from the storage tanks, both at $32/bbl of oil.

Source: DRI estimates based on information provided by SWEC.

TABLE 5.1-17. TOTAL HYDROCARBON EMISSIONS FROM THE PLANT

Stream Number	Emission Source	Control Description	Hydrocarbon Emissions (lb/hr)
24	Diesel Equipment	Catalytic Converters	12.7
31	Flue Gas Discharge System	--	6,261.7*
112	Valves, Pumps, etc.	Maintenance	36.5
44, 47	Product Storage	Floating Roof Storage Tanks	65.5
TOTAL			6,376.4

* According to the information from Rio Blanco Oil Shale Co., February 1981, about 232 lb/hr of hydrocarbons (expressed as CH_4) are estimated from the 4,400 TPSD Lurgi module (the processed shale rate is 3,518 TPSD). The reported value is extrapolated for the commercial operation (94,956 TPSD of processed shale).

Source: SWEC estimates, except as noted.

The easiest and most cost-effective way to control CO emissions is to use excess oxygen in the combustion processes to ensure complete combustion. When incomplete combustion does occur, catalytic converters or thermal or chemical oxidizers may be used to oxidize the remaining CO to CO_2.

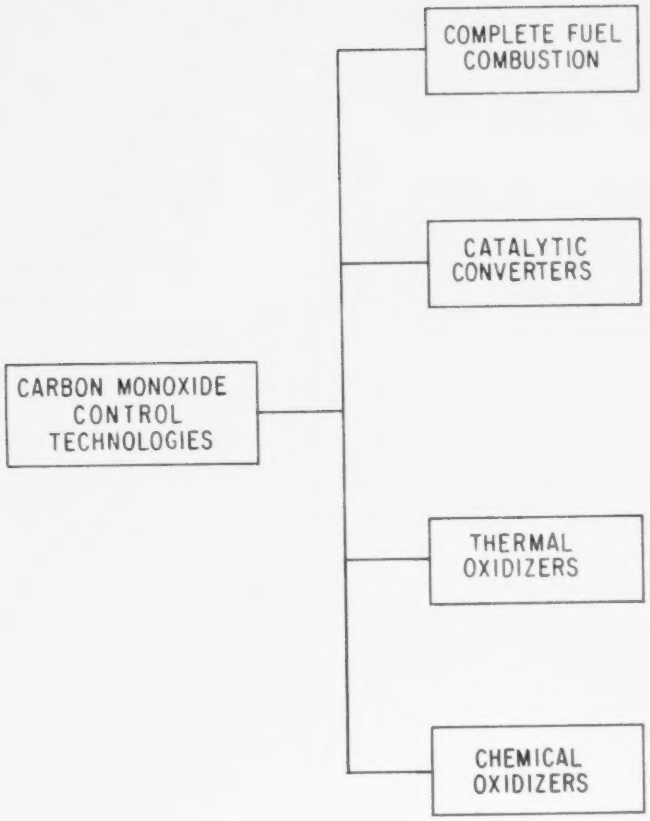
Inventory of Control Technologies--

Figure 5.1-9 shows a list of the applicable carbon monoxide control technologies, and Table 5.1-18 describes in detail the features of these control methods.

Complete fuel combustion controls CO emissions by not allowing them to be formed. This is done by operating with enough excess air to ensure complete oxidation of all carbon to CO_2 instead of only partial oxidation to CO. When CO is formed in a combustion process, a catalytic converter or thermal or chemical oxidizer can be used.

Carbon Monoxide Control Technologies Analyzed--

By far, the largest amount of CO is emitted from the Lurgi flue gas discharge system. The sources of this CO may be the incomplete combustion of the residual organic matter on the processed shale, decomposition of the carbonate minerals, and a steam/coke reaction in the processed shale quencher/moisturizer. To maximize the combustion of the organic residue, an


CARBON MONOXIDE CONTROL TECHNOLOGIES
COMPLETE FUEL COMBUSTION
CATALYTIC CONVERTERS
THERMAL OXIDIZERS
CHEMICAL OXIDIZERS

SOURCE: SWEC

FIGURE 5.1-9 CARBON MONOXIDE CONTROL TECHNOLOGIES

TABLE 5.1-18. KEY FEATURES OF CARBON MONOXIDE CONTROL TECHNOLOGIES

Control Technology	Operating Principle	Performance	Development Status	Advantages	Disadvantages
Complete Fuel Combustion	Combustion process is operated with excess air to ensure complete oxidation of all carbon to CO_2, instead of only partial oxidation to CO.	Can convert close to 100% of all carbon in the fuel to CO_2.	Commercially proven.	Eliminates the need for downstream equipment to complete the conversion of CO to CO_2.	An adequate air:fuel ratio must be maintained.
Catalytic Converters	Hot exhaust gas is passed over a catalyst where the CO in the gas is reacted with the excess air in the exhaust gas and is converted to CO_2.	Can convert up to 90% of all CO in diesel exhaust gas to CO_2; for other fuel burning processes, up to 99% conversion is possible.	Commercially proven.	Does not require any fuel and has no moving parts so that routine maintenance is minimal.	The catalyst, which is expensive, must be replaced periodically.
Thermal Oxidizers	Waste gas streams containing CO are burned with excess air and additional fuel if needed to completely oxidize all CO to CO_2.	Can convert up to 100% of all CO in the gas stream to CO_2.	Commercially proven.	Will ensure complete oxidation of CO to CO_2 and complete oxidation of any other unwanted components in the gas stream.	Can have a high energy requirement when supplemental fuel is used.
Chemical Oxidizers	Gas streams containing CO are scrubbed with a solution containing a chemical oxidizing agent which oxidizes the CO to CO_2.	Can convert up to 99% of all CO in the gas stream to CO_2.	Commercially proven.	Oxidizes the CO to CO_2 without using fuel to heat up the entire gas stream.	Requires the use of expensive chemicals.

Source: SWEC based on information from Research and Education Association, 1980.

excess of air is used. Decomposition of carbonates is unavoidable because the processed shale recycle stream has to attain a high temperature to provide the heat of retorting. The steam/coke reactions may also be unavoidable.

The CO content of the flue gas is reported to be less than 90 ppmv. This may be reduced further by the post-combustion of the flue gas; however, due to the large volume and low heating value of the flue gas, it would be impractical.

Diesel-powered equipment is another source of the CO emissions. The diesel engines are equipped with catalytic converters to control the CO. Since the converters also control hydrocarbons, they have been included under hydrocarbon emission control.

Total Carbon Monoxide Emissions--

Table 5.1-19 summarizes the carbon monoxide emission sources and control equipment used for the emissions.

TABLE 5.1-19. TOTAL CO EMISSIONS FROM THE PLANT

Stream Number	Emission Source	Control Description	CO Emissions (lb/hr)
24	Diesel Equipment	Catalytic Converters	34.8
31	Lurgi Flue Gas Discharge System	--	657.4*
TOTAL			692.2

* According to the information from Rio Blanco Oil Shale Co., February 1981, the flue gas contains about 90 ppmv CO.

Source: SWEC estimates, except as noted.

5.1.6 Control of Other Criteria Pollutants

In addition to the primary air pollutants discussed so far, there may be other criteria pollutants, such as lead, mercury, beryllium and fluorides, emitted from the Lurgi-Open Pit facility. Some of these pollutants are nonvolatile; therefore, they may be released only as fugitive dust constituents. Any control of the dust will also serve to control the nonvolatile pollutants. Volatile pollutants may potentially be released with the Lurgi flue gas and/or the tail gas from the Stretford process. Some pollutants do not occur naturally and some are unlikely to form during oil shale processing.

5.1.7 Control of Noncriteria Air Pollutants

Meaningful test data are not available to determine whether emissions of noncriteria air pollutants are a concern. Consequently, no information on control technologies for such pollutants was generated for this manual. Mention of species such as POMs (U.S. EPA, 1980) and trace elements such as arsenic (Fox, Mason and Duvall, 1979; Girvin, Hadeishi and Fox, June 1980) are noted.

5.2 WATER MANAGEMENT AND POLLUTION CONTROL

As in other industries and oil shale operations, the Lurgi-Open Pit plant--from mining activities to final product and waste disposition--will produce water effluents which will require proper disposal. These effluents may contain the following pollutants:

- Suspended Matter, Oil and Grease
- Dissolved Gases and Volatiles
- Dissolved Inorganics
- Dissolved Organics.

This section describes the current, commercially available alternate systems for controlling the above pollutants. The following subsections provide inventories of control technologies for each of the pollutant classes, a discussion of advantages and disadvantages, and important points to consider in selecting a particular technology. The performance, design, and cost data for the leading technologies are also presented.

5.2.1 Suspended Matter, Oil and Grease

Undissolved matter found in wastewater effluents includes solid particles as well as oils and greases. The solids are usually the raw and processed shale particles that are washed into the retort water and those that are entrained in the retort vapors and subsequently removed in the gas condensates. The retort water and gas condensate also contain trapped oil and oil-in-water emulsions. Service and storm runoffs contain suspended matter, as well as oils and greases. Also, the source water contains suspended soil particles and debris.

In general, the control of suspended matter at oil shale plants will be accomplished using conventional technology. For example, clarification in gravity settlers (with addition of flocculants) and multimedia filtration will, in most cases, provide adequate control. Associated energy consumption and costs are generally low.

The control of undissolved oils and greases in oil shale wastewaters has not been studied in detail. API-type gravity settlers have the potential to provide adequate control for most of the waste streams generated. It is possible, however, that some wastewaters will contain oil-in-water emulsions; if so, additional control steps may be required. Heating the water or adding chemicals may be sufficient to break the emulsion; otherwise, filter coalescence (or possibly ultrafiltration) may be required.

The degree to which emulsified oil needs removal is dependent on downstream processing and reuse. In cooling towers, the oil may foul heat exchange surfaces and thus require prior removal. Similarly, fouling, and possibly foaming, may occur when stripping the retort water or gas condensate. The extent to which such problems will arise is not known.

The energy consumption and cost of oil separation by gravity means are generally low. Thermal or chemical treatment, if required, would cause some increase in costs. Filter coalescence and, in particular, ultrafiltration generally are more costly and would be considered only if other procedures prove inadequate.

Inventory of Control Technologies--

Figure 5.2-1 shows different types of technologies that apply to control of suspended matter and oils and greases. Key features of these technologies are provided in Table 5.2-1.

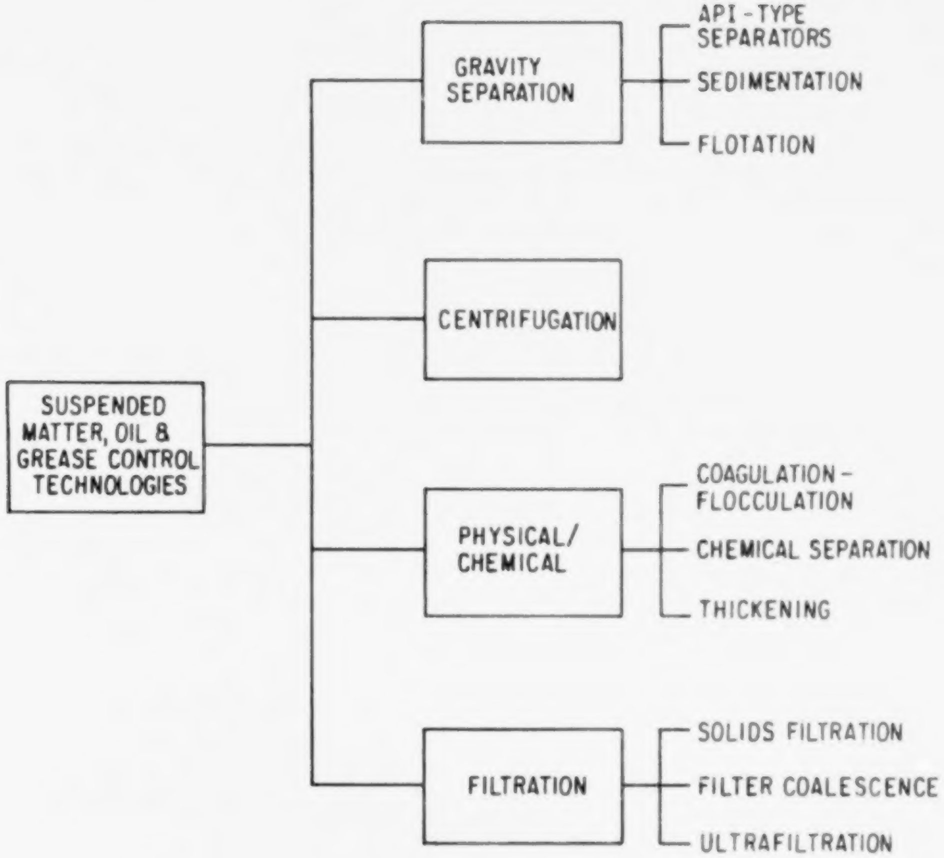
API-type separators. For gravity separation of oil in large holding tanks, separators should be designed within the following limits: (a) horizontal velocity of less than 3 fpm, (b) depth between 3-8 ft, and (c) depth-to-width ratio of approximately 0.4. Oil is skimmed from the surface and collected for reuse or disposal. Gravity separation is not effective for emulsified oils that might be present in some retort waters (American Petroleum Institute, 1969).

Sedimentation. This is a gravity process in which the solid phase settles and is withdrawn as a slurry. Clarification may be carried out in large holding ponds, plate (lamella) settlers or hydrocyclones. Chemicals (flocculants and coagulants) may be added to precipitate salts (softening) or to aid settling of suspended solids (Humenick, 1977).

Flotation. This is a gravity process in which the solid phase rises to the surface and is skimmed off as a slurry. Air bubbles may be introduced into the flotation vessels to assist separation (Humenick, 1977).

Centrifugation. This is a modified gravity method to afford separation or settling of fine, suspended matter and oils. The wastewater is subjected to a radial force greater than the gravity field by rapidly rotating it. Suspended matter denser than water moves radially away from the center of rotation, while the lighter matter moves toward the center. Concentrated matter can be removed periodically or in a continuous manner. For continuous operations, the sludge should be fluid to facilitate its removal. The technology may not be applicable to highly viscous fluids.

Coagulation - flocculation. Fine particles suspended in a fluid are subjected to size enlargement by addition of chemicals (coagulants and flocculants), then allowed to settle by gravity or under applied force. Gentle agitation alone sometimes may afford the flocculation of the particles. The


SUSPENDED MATTER, OIL & GREASE CONTROL TECHNOLOGIES
GRAVITY SEPARATION
API - TYPE SEPARATORS
SEDIMENTATION
FLOTATION
CENTRIFUGATION
PHYSICAL/ CHEMICAL
COAGULATION - FLOCCULATION
CHEMICAL SEPARATION
THICKENING
FILTRATION
SOLIDS FILTRATION
FILTER COALESCENCE
ULTRAFILTRATION

SOURCE: WPA

FIGURE 5.2-1 SUSPENDED MATTER, OIL AND GREASE CONTROL TECHNOLOGIES

TABLE 5.2-1. KEY FEATURES OF CONTROL TECHNOLOGIES FOR SUSPENDED MATTER, OILS AND GREASES

Control Technology	Operating Principle	Components Removed	Removal Efficiency	Feed Requirements/ Restrictions	By-products and Wastes	Comments
Gravity Separation (API-type Separators, Sedimentation, Flotation)	Provision of adequate residence time in a stagnant vessel to allow suspended matter to separate into lighter and heavier than water components. Extra surfaces may be included to save space (lamella settler), or rising air bubbles may be used to assist separation (dissolved air flotation).	Suspended solids, tars, oils, immiscible liquids.	90+% removal of TSS typical for sedimentation, 50% for flotation.	Minimum feed stream turbulence.	Oils, sludges, solids.	Not useful for emulsions or very fine particles.
Centrifugation	A greater than gravity force field is applied by rapid rotation to accelerate the separation.	As above.	As above.		As above.	More expensive than gravity separation. Used for predrying sludges from gravity separation devices or for separation of fine particles.
Physical/Chemical (Coagulation-Flocculation, Chemical Separation, Thickening)	Use of agents to promote the coalescence of fine suspended solids, tars and oils. Generally used in conjunction with a gravity separation process.	Promotes removal of finely dispersed particles.	90+% removal of fine solids is achievable with proper design.	A wide range of commercial flocculants are available.	Same as gravity separation.	Widely used in makeup water treatment systems to remove fine solids.
Filtration (Solids Filtration, Filter Coalescence, Ultrafiltration)	Involves passing wastewater through a suitable filter medium. Filter material is discarded or cleaned by backflushing.	Depends on medium. Both coarse and fine structure materials are used industrially.	90-99% is typical.	Filter media (sand,clay, fabric or polymeric membrane).	Filter backwash, spent filter media.	Filter coalescence or ultrafiltration are useful for oil emulsions

Source: WPA.

technology may also be applicable to liquid dispersions and liquid particulates.

Chemical separation. Addition of chemicals to break emulsion may be used in conjunction with filtration and is normally followed by gravity separation. The type and dosage of chemicals required is determined by trial (American Petroleum Institute, 1969). Chemicals may also be added to precipitate salts or to increase crystal size.

Thickening. Slurries previously obtained from gravity, centrifugation, and filtration methods can be further concentrated, or thickened, by addition of chemical agents or binders. The thickened slurry may then be subjected to the same methods for final disposition (Adams and Eckenfelder, 1974; Humenick, 1977).

Solids filtration. The water stream is passed through a filter medium which holds back the solid phase. Filters may be of the fabric type, as in plate and frame, rotating drum (vacuum) and cartridge units, or granular, as in sand filters. Filtration is generally more expensive than sedimentation but can remove smaller particles (Humenick, 1977).

Filter coalescence. Gravity separation of oil from water is standard industrial and refinery practice; however, the API-type separators are inadequate for very small oil particles. One very important method for removal of small oil droplets is coalescence (Water Purification Associates, December 1975).

When a dispersion of micron-sized droplets of one liquid (oil) in another (water) flows through an appropriate porous solid, coalescence of the dispersed phase is induced and separation of the liquids results. The dispersed phase can be allowed to accumulate without leaving the porous medium, with periodic regeneration to remove accumulated oil.

Filter media are usually either the packed fibrous type (e.g., fiber glass, steel wool) or unconsolidated granular materials (e.g., sand, gravel, crushed coal). Because of their large specific surface and high voids, fibrous media are usually more efficient in removing droplets for a given bed depth than are granular media. However, fibrous media are more susceptible to blockage by suspended solids and are more difficult to regenerate, in addition to being more costly than most granular media.

Advantages of filter-coalescers include high separation efficiency for dilute suspensions of very small droplets, potentially small space requirements, the possibility of continuous operation, and the potential for the recovery of the dispersed phase. Disadvantages of this process are that suspended solids can accumulate to require frequent medium regeneration or replacement, and pumping costs can be substantial. As far as is known, the system has not been evaluated on retort waters, and extensive pilot plant testing would be required to determine its feasibility on these waters.

Ultrafiltration. Passage through a submicron-sized membrane filter separates emulsified oil as well as suspended matter and large organic molecules (MWt ≥ 1,000). The oil droplets are collected in the concentrate and removed by gravity separation. This process is significantly more costly than normal filtration (Water Purification Associates, December 1975).

Control Technologies Analyzed--

The streams that require removal of suspended matter, oils and greases are:

- Mine Water (stream 4)
- Gas Liquor (stream 41)
- Runoff and Leachates (streams 92, 93)
- Blowdowns and Concentrates (streams 88, 104).

Mine water is obtained from dewatering of the deep aquifers under Tract C-a. While the water does not contain any oils and greases, it does contain suspended matter. Sedimentation by gravity settling and clarification with addition of alum are the approaches proposed to reduce the suspended matter in the mine water. Table 5.2-2 presents the design features and cost data for clarification, and Figure 5.2-2 shows a cost curve for the clarifier. This activity could be considered as part of the process rather than pollution control.

In the Lurgi retorting process, gas liquor is condensed along with light oils in the third condensation tower. It may also contain some particulate matter that was not removed in the cyclones and two previous towers, but this is not envisioned as a problem; however, the gas liquor will need separation from the light oils. An API-type oil/water separator with channel covers was examined for this purpose. As stated earlier, the separators are not fully established as useful devices for shale oils, but difficulty in achieving separation from light oils is not anticipated. Table 5.2-3 and Figure 5.2-3 present the cost and design information and the cost curve, respectively, for the API separator.

Service and fire water runoff, storm runoff, and leachate from shale piles may contain oily materials. Again, an API-type oil/water separator was examined as the control. This will also allow separation of suspended matter along with the water. The cost and design data for this separator are given in Table 5.2-4, while a cost curve is already included in Figure 5.2-3.

The blowdowns, sludges, and concentrates from various processing units will also contain suspended matter. These streams are collected in an equalization pond for possible use in processed shale moisturizing. Since gravity settlement affords separation of the suspended matter, the equalization pond also might be viewed as a pollution control. Its design and cost are presented in Table 5.2-5, and a cost curve is given in Figure 5.2-4.

Other Technologies Analyzed--

In the event that the excess mine water is reinjected into the aquifer (instead of discharging it on the surface), even more water will need to be

TABLE 5.2-2. DESIGN AND COST OF MINE WATER CLARIFICATION[a]

Item	Unit	Quantity
Mine Water Flow Rate	gpm	16,500
Flow Rate/Clarifier	gpm	970
Number of Clarifiers	--	17
Diameter	ft	40
Area of Clarifier[b]	10^3 ft^2	22.3
Alum Rate (30 ppm)	ton/yr	980
Fixed Capital Cost	\$$10^3$	2,560
Direct Annual Operating Cost	\$$10^3$	
Maintenance @ 4%[c]		84
Alum @ 12¢/lb		235
TOTAL		319
Total Annual Control Cost[d]	\$$10^3$	961

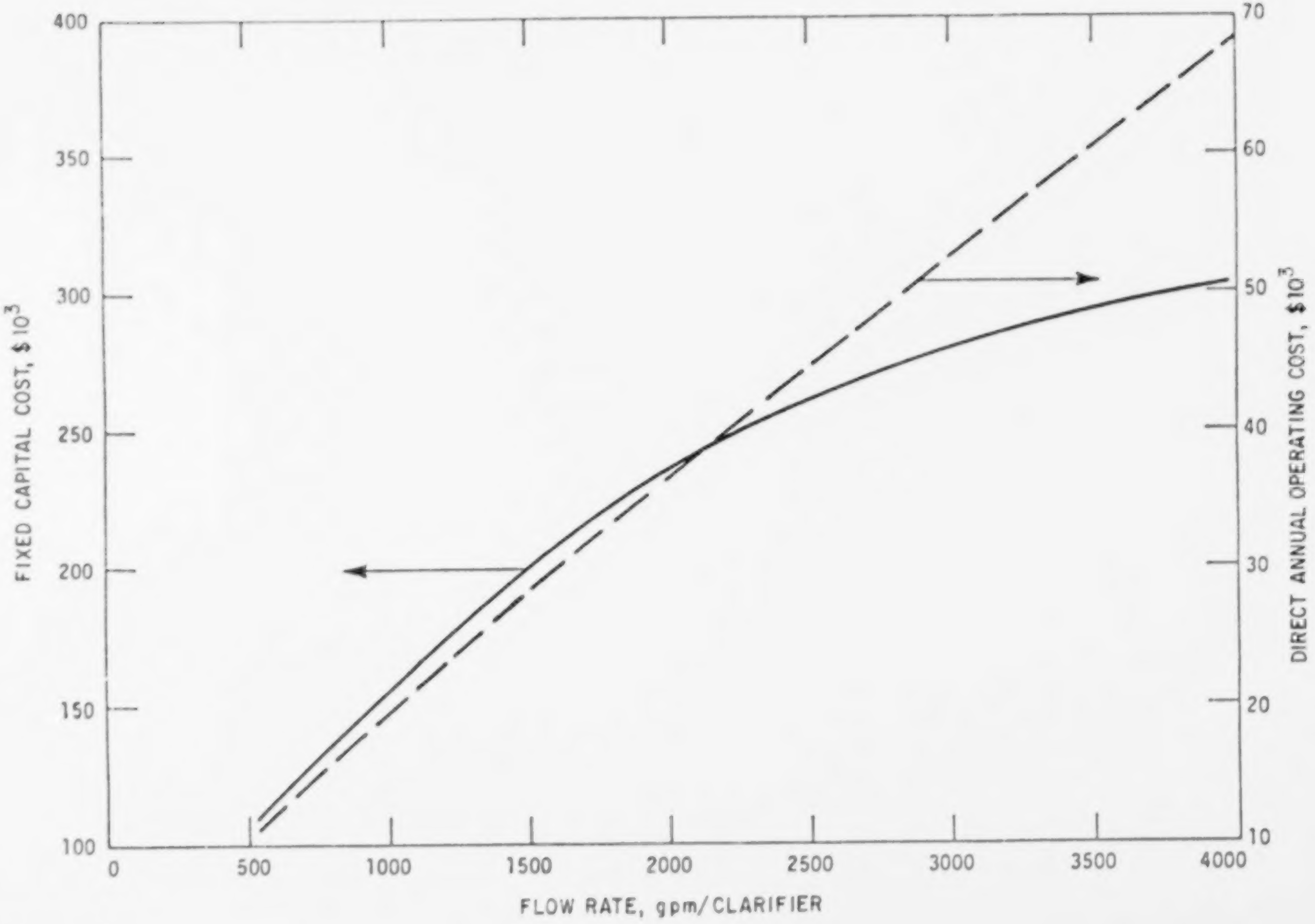
[a] This technology could be considered as part of the process rather than pollution control.

[b] Retention time and rise rate are 120 min. and 1 gpm/ft^2, respectively.

[c] Maintenance is based on the fixed capital cost less contingency.

[d] See Section 6 for details on computation of the total annual control cost.

Source: WPA estimates.



SOURCE: WPA

FIGURE 5.2-2 COST OF MINE WATER CLARIFICATION

TABLE 5.2-3. DESIGN AND COST OF API OIL/WATER SEPARATOR FOR GAS LIQUOR

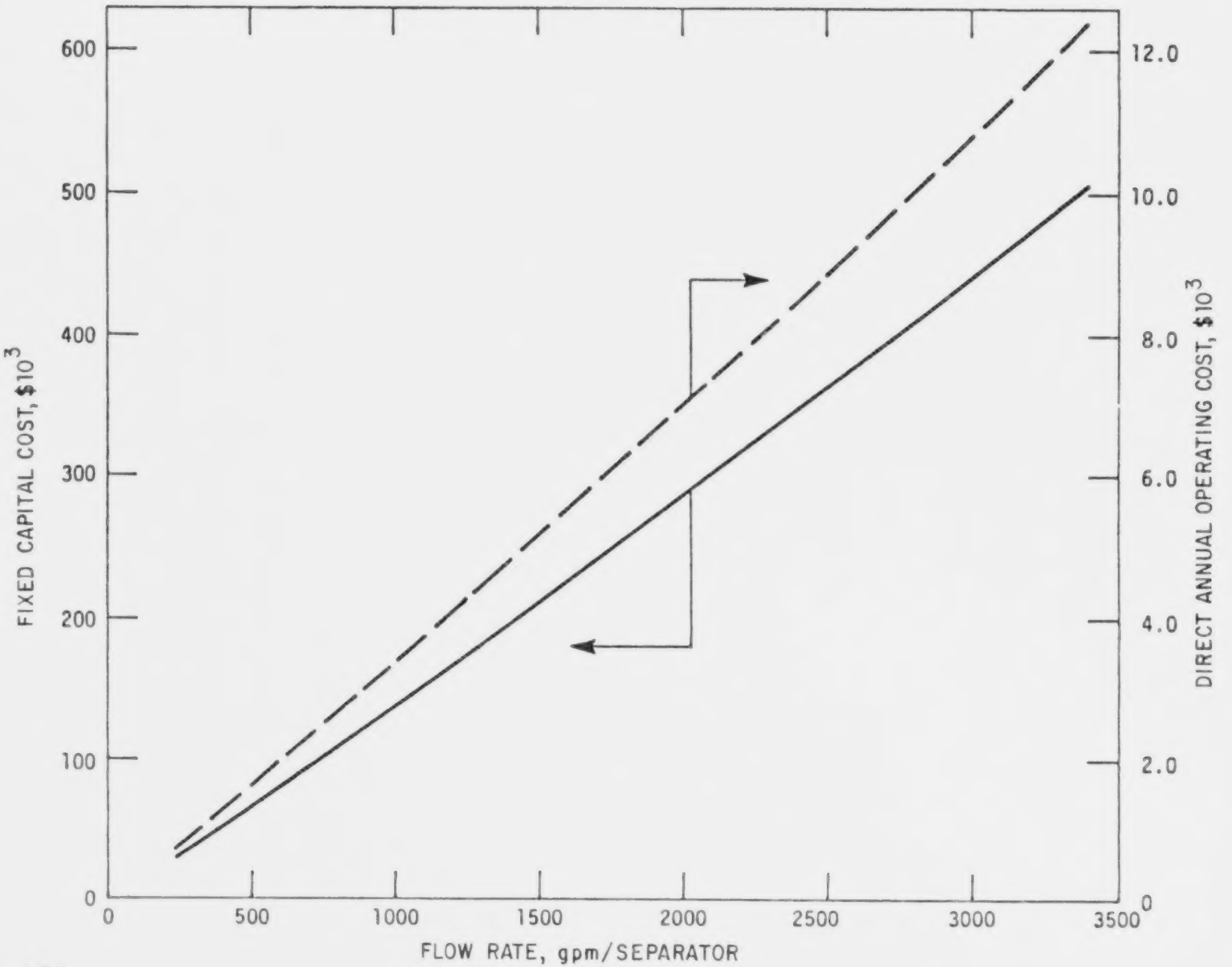
Item	Unit	Quantity
Gas Liquor Flow Rate	gpm	586
No. of Channels (1 standby)	--	2
Channel Cross Sectional Area	ft^2	27
Channel Depth	ft	6.5
Channel Length	ft	50
Fixed Capital Cost[a]	$\$10^3$	161
Direct Annual Operating Cost[a]	$\$10^3$	
Maintenance @ 3%[b]		4
Total Annual Control Cost[c]	$\$10^3$	35

[a] The fixed capital cost and direct annual operating cost for the standby channel are included.

[b] Maintenance is based on the fixed capital cost less contingency.

[c] See Section 6 for details on computation of the total annual control cost.

Source: WPA estimates based on information from American Petroleum Institute, 1969.



SOURCE: WPA

FIGURE 5.2-3 COST OF OIL/WATER SEPARATION

TABLE 5.2-4. DESIGN AND COST OF OIL/WATER SEPARATOR FOR RUNOFFS AND LEACHATE

Item	Unit	Quantity
Runoff Flow Rate	gpm	169
No. of Channels (1 standby)	--	2
Channel Cross Sectional Area	ft^2	8
Channel Depth	ft	3
Channel Length	ft	50
Fixed Capital Cost[a]	$\$10^3$	41
Direct Annual Operating Cost[a]	$\$10^3$	
Maintenance @ 3%[b]		1
Total Annual Control Cost[c]	$\$10^3$	11

[a] The fixed capital cost and direct annual operating cost for the standby channel are included.

[b] Maintenance is based on the fixed capital cost less contingency.

[c] See Section 6 for details on computation of the total annual control cost.

Source: WPA estimates based on information from American Petroleum Institute, 1969.

TABLE 5.2-5. DESIGN AND COST OF EQUALIZATION POND

Item	Unit	Quantity
Total Water Flow Rate	gpm (acre-ft/yr)	2,525 (4,065)
Pond Area	acre	3.27
Pond Depth	ft	10
Fixed Capital Cost	$\$10^3$	181
Direct Annual Operating Cost	$\$10^3$	
Maintenance @ 2%[a]		3
Total Annual Control Cost[b]	$\$10^3$	46

[a] Maintenance is based on the fixed capital cost less contingency.

[b] See Section 6 for details on computation of the total annual control cost.

Source: WPA estimates.

dewatered because some of the reinjected water will flow back to the dewatering wells.

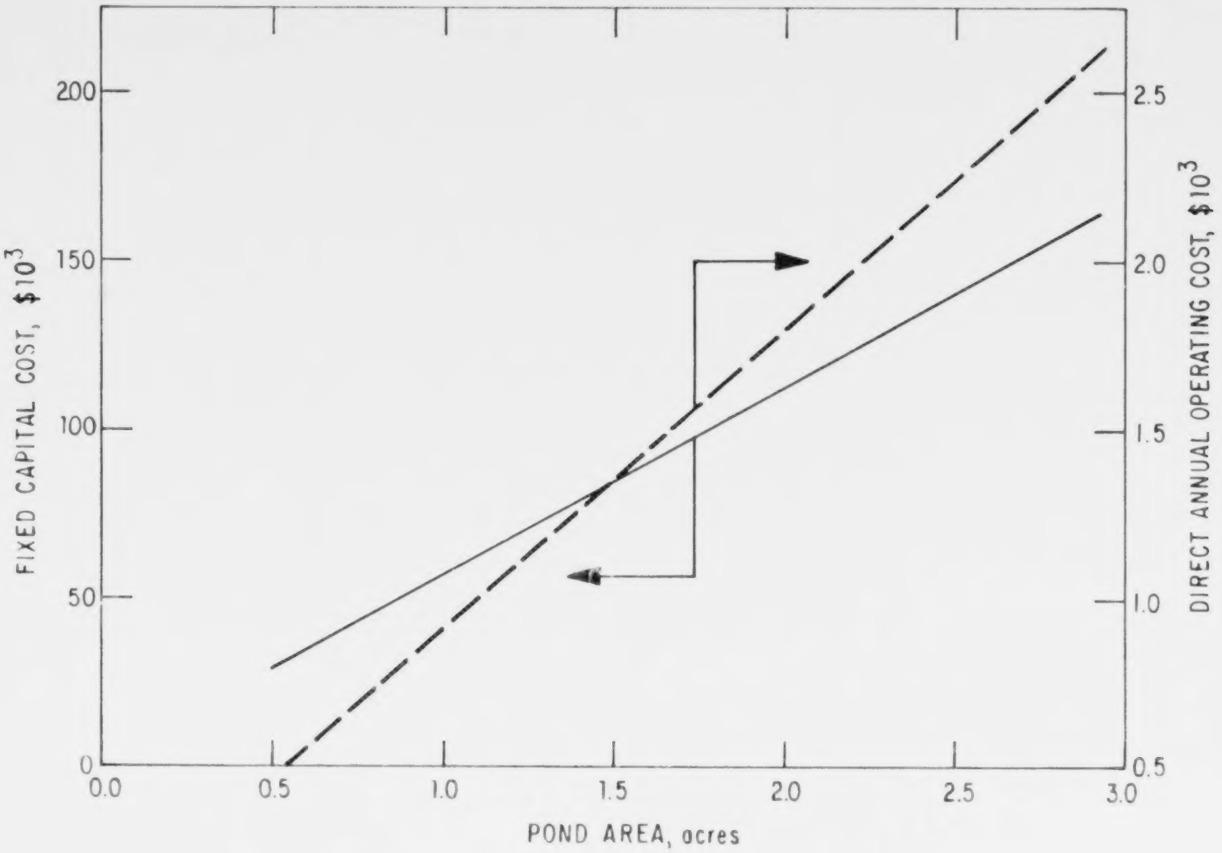
The water to be reinjected would need to be clarified. This should be performed in closed clarifiers to avoid exposure of the excess water to the environment. A closed clarifier was examined for the reinjection water, and its design and cost information is presented in Table 5.2-6. A cost curve based on the design of the clarifier is shown in Figure 5.2-5.

5.2.2 Dissolved Gases and Volatiles

Dissolved gases include ammonia, carbon dioxide, and hydrogen sulfide, while volatile materials are low molecular weight organics. Methods for removing these substances from water are summarized in Figure 5.2-6. Steam stripping is the most likely process to be used and has been successfully demonstrated on a laboratory scale for some oil shale wastewaters (Hicks and Liang, January 1981).

Inventory of Control Technologies--

Table 5.2-7 presents an inventory of applicable control technologies, along with their key features, for the dissolved volatiles. Basically, most technologies involve stripping of the dissolved gases by either elevating the temperature, applying vacuum, or displacement with carrier gases. More specific removal can be accomplished by using an adsorbent selective for the gas in question.



SOURCE: WPA

FIGURE 5.2-4 COST OF EQUALIZATION POND

TABLE 5.2-6. DESIGN AND COST OF EXCESS MINE WATER CLARIFICATION[a]

Item	Unit	Quantity
Excess Mine Water Flow Rate	gpm	15,330
Flow Rate/Clarifier	gpm	510
Number of Clarifiers	--	30
Diameter	ft	30
Area of Clarifier	10^3 ft^2	20.7
Alum Rate (30 ppm)	ton/yr	900
Fixed Capital Cost	$\$10^3$	3,545
Direct Annual Operating Cost	$\$10^3$	
Maintenance @ 4%[b]		115
Alum @ 12¢/lb		216
TOTAL		331

[a] This technology could be considered as part of the process rather than pollution control.

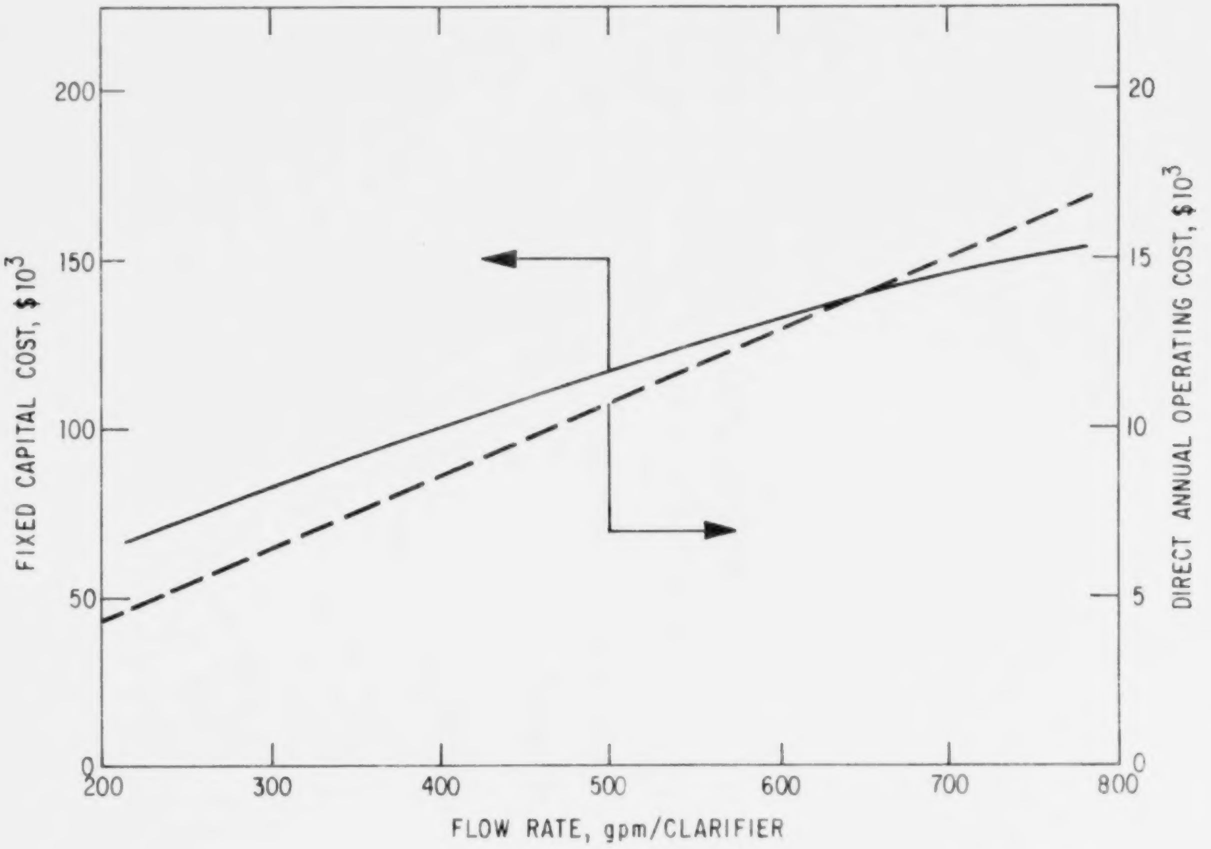
[b] Maintenance is based on the fixed capital cost less contingency.

Source: WPA estimates.

Steam stripping. Steam stripping of sour waters (e.g., waters containing dissolved ammonia and hydrogen sulfide) and coke-oven liquors (e.g., waters containing dissolved ammonia and carbon dioxide) is standard practice in the petroleum and steel industries. Stripping has also been used as part of the "Phenosolvan" process on coal gasification process condensates (American Petroleum Institute, March 1978; Beychok, 1967).

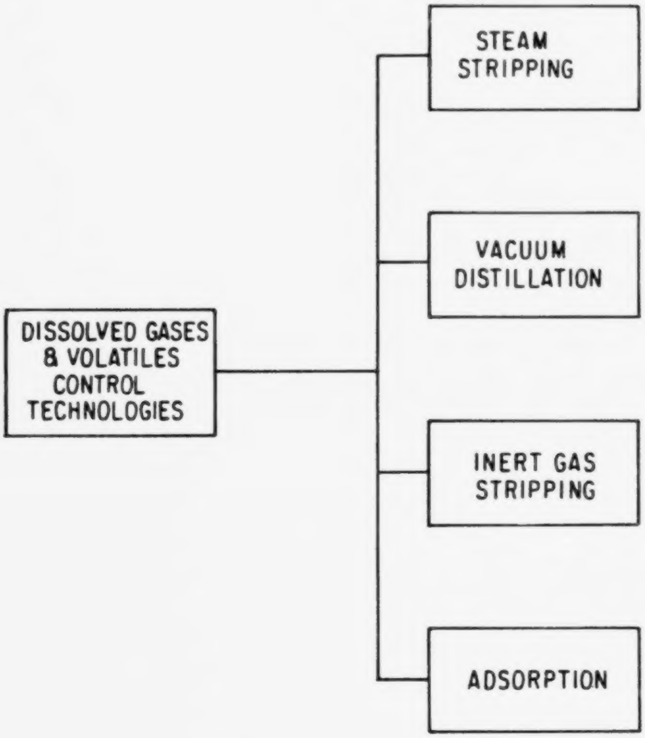
The dissolved gases are stripped from the solution by bubbling steam through it, generally in packed or tray columns. The steam may be directly sparged (live) or used indirectly in a reboiler, as in distillation columns. The stripped gases, along with other volatile materials, are removed in a relatively concentrated gas stream which may be treated for adsorption/recovery of a specific substance or incinerated. Carbon dioxide is readily stripped at efficiencies of +99%; ammonia strips less easily, and pH elevation may be required in some cases for 99% removal. Hydrogen sulfide does not strip as easily as carbon dioxide but can generally be removed down to the 10-20 ppm range. Costs are for equipment and steam and are proportional to the volume of water to be treated.

Steam requirements range from approximately 10 to 15 lbs steam per 100 lbs water treated. For a given separation, a greater column height is



SOURCE: WPA

FIGURE 5.2-5 COST OF EXCESS MINE WATER CLARIFICATION


DISSOLVED GASES & VOLATILES CONTROL TECHNOLOGIES
STEAM STRIPPING
VACUUM DISTILLATION
INERT GAS STRIPPING
ADSORPTION

SOURCE: WPA

FIGURE 5.2-6 DISSOLVED GASES AND VOLATILES CONTROL TECHNOLOGIES

TABLE 5.2-7. KEY FEATURES OF CONTROL TECHNOLOGIES FOR DISSOLVED GASES AND VOLATILES

Control Technology	Operating Principle	Components Removed	Removal Efficiency	Feed Requirements/ Restrictions	By-products and Wastes	Comments
Steam Stripping	Increasing temperature and providing a positive flow of steam through the wastewater. Removes volatile organics and inorganics with overhead steam.	NH_3, acid gases (CO_2, H_2S, HCN), light hydrocarbons.	90+% of "free" ammonia and acid gases typical. Hydrocarbon removal varies with volatility of stripped components.	Acid/caustic for pH adjustment optional.	Stripped gases, uncondensed steam.	Acid/caustic addition can be used to improve the efficiency and selectivity of the stripping process.
Vacuum Distillation	Low pressure, low temperature stripping process.	As above.	As above.	As above.	Stripped gases.	High energy requirements. Not cost competitive in a plant where stripping steam is readily available.
Inert Gas Stripping	Same as stream stripping, but using air, nitrogen, or other available inert gas in place of steam.	As above.	As above.	As above.	Stripped and inert gases.	Normally used at ambient temperature; most suitable for low concentration wastes.
Adsorption	Adsorption of NH_3 onto clinoptilolite and volatile organics onto polymeric resins.	NH_3, volatile organics.	High removal efficiencies possible.	Not suitable for high concentrations.	Regenerant and adsorbent wastes.	Generally used as a polishing step.

Source: WPA.

required for a lower steam rate. The selection of steam rate and column height is based on energy and equipment costs.

The stripped gases may be incinerated or treated further to recover ammonia and sulfur. Ammonia may be recovered as anhydrous ammonia, aqua (20-30%) ammonia or ammonium sulfate. In cases where the sulfate is derived from flue gas desulfurization, the sulfate route may be viable depending, in part, on the marketability of ammonium sulfate and on the costs of alternative flue gas desulfurization processes. Because oil shale plants generally will have ammonia available as a by-product, SO_2 scrubbing with NH_3 may be attractive when the technology is sufficiently developed and tested. Recovery of anhydrous ammonia involves considerable capital and energy (steam) requirements, but these are partially offset with by-product ammonia sales. The stability of the ammonia market must be considered when selecting a recovery process.

Vacuum distillation. Distillation at reduced pressure has many industrial applications, but these primarily involve distillation or fractionation of compounds with high boiling points or low thermal stability. The method may be applicable to stripping of gases and volatile compounds, but the energy requirements are high relative to those for steam or inert gas stripping.

Inert gas stripping. This method is applicable to dilute, or low strength, wastewaters for which steam stripping may not be practical. The operating principle is similar to that for steam stripping, except air, nitrogen, carbon dioxide, or other inert gases may be used. Its application to high strength liquids is generally not practical because large column heights and gas compression costs are required.

Adsorption. Dissolved gases and volatile components may be adsorbed on specific surface-active materials by passing wastewaters through a bed of the adsorbent. The gases may then be desorbed thermally, and the regenerated adsorbent is recycled. This method is generally used in trace removal applications.

Control Technologies Analyzed--

The streams that may require removal of dissolved gases and volatiles are:

- Gas Liquor (stream 41)
- Compression Condensate (stream 49).

The compression condensate is also a retort gas condensate obtained during compression and cooling of the retort gas; therefore, it is combined with the gas liquor for treatment. The condensates are previously freed from oil and emulsion in the oil/water separator, but some polar organics, such as phenols and fatty acids, remain dissolved. A portion of the dissolved organics can be steam stripped along with other dissolved gases. The gas liquor also contains a significant amount of ammonia, both free as well as fixed, with sulfur dioxide and carbon dioxide. Since the intended use of the gas liquor is in processed shale moisturizing, most of the fixed ammonia must also be removed from the gas liquor prior to its use; otherwise, it may be released into the environment upon contact with the alkaline processed shale. Steam stripping alone would remove free ammonia and other volatile components from the liquor, but pH adjustment may be necessary to release fixed ammonia. A further control of the released ammonia is also desirable and this may be accomplished with an ammonia recovery plant.

Ammonia recovery was examined as a control for the gas liquor. The design specifications for the ammonia recovery plant are given in Table 5.2-8, the cost is presented in Table 5.2-9, and a cost curve is presented in Figure 5.2-7. The description and material balance for the process are presented in Sections 3.3.8 and 4.2.7, respectively.

5.2.3 Dissolved Inorganics

Dissolved inorganics are usually not a problem unless the compounds are judged to be hazardous (e.g., trace metals) or when fouling of equipment (e.g., boilers) occurs because of the high salt content of the waters being used. Natural waters and waters that come into contact with the solids may need to be treated if they are intended for critical uses in the plant. Processed shale moisturizing, on the other hand, may not require control of dissolved inorganics. In fact, waters with high salt content can be used for this purpose, thereby avoiding the need for other controls. Since gas condensates do not contain significant amounts of dissolved inorganics, a treatment may not be necessary.

Inventory of Control Technologies--

Methods for removal of dissolved inorganics are shown in Figure 5.2-8, while some of the key features of the technologies are presented in Table 5.2-10. The operating principles for some of the methods shown in the figure are detailed below.

Precipitation. Chemicals may be added to precipitate salts, e.g., lime addition for carbonate (hardness) removal. Processed shale is also believed to behave like a softener for inorganic carbon reduction (Humenick, 1977). The process is simple, but it will usually require the use of other methods (e.g., gravity separation, centrifugation, filtration) to remove the precipitate.

Ion exchange. Cations and anions in solution are replaced with hydrogen and hydroxyl ions on exchange resins capable of producing a water virtually free of common salts. The resins are regenerated with relatively strong acid and alkali solutions, and the regenerant wastes must be controlled. Costs go up with increasing concentration of salts in the water. Ion exchange is normally used only where a very clean water is required from a relatively clean or mildly brackish supply. The organics present are not removed and may foul the exchange resins (Calmon and Gold, 1979).

TABLE 5.2-8. DESIGN OF AMMONIA RECOVERY SYSTEM*

Design Parameter	Unit	Quantity
Gas Condensate Feed Rate to Stripper Column	gpm	594
Ammonia Rate	lb/hr	1,883
Steam Rate	10^3 lb/hr	53
Cooling Water Circulated	gpm	1,080
Electricity	kW	47
Chemicals		
H_3PO_4	lb/hr	13
NaOH	lb/hr	293
Steam Stripping Column		
Diameter	ft	6.3
Height	ft	95
Material	--	CS/SS
Reboilers on Steam Stripping Column		
Number	--	1
Surface area (each)	ft^2	2,300
Material	--	CS/SS
Heat Exchanger on Steam Stripping Column		
Number	--	3
Surface area (each)	ft^2	5,000
Material	--	CS/SS
Absorption Column		
Diameter	ft	5
Height	ft	50
Material	--	SS
Reboiler on Absorber		
Surface area	ft^2	701
Material	--	CS/SS
Heat Exchanger on Absorber		
Surface area	ft^2	948
Material	--	CS/SS
Stripper Tower		
Diameter	ft	3.3
Height	ft	60
Material	--	SS

(Continued)

TABLE 5.2-8 (cont.)

Design Parameter	Unit	Quantity
Heat Exchanger on Stripper		
Surface area	ft^2	1,137
Material	--	SS
Fractionator		
Diameter	ft	1.5
Height	ft	64
Material	--	SS
Fractionator Feed Tank		
Diameter	ft	7
Height	ft	4.3
Capacity	gal	1,278
Material	--	SS
Reboiler on Fractionator		
Surface area	ft^2	209
Material	--	CS/SS
Heat Exchanger on Fractionator		
Surface area	ft^2	645
Material	--	CS/SS
Flash Drum		
Diameter	ft	4
Height	ft	1.4
Capacity	gal	142
Material	--	SS
Lean Solution Cooler		
Surface area	ft^2	1,554
Material	--	CS/SS
Solution Heat Exchanger		
Surface area	ft^2	303
Material	--	SS

* This table is based on the Phosam-W process, which is only one example of many available processes for the recovery of ammonia.

Source: WPA estimates based on information provided by U.S.S. Engineers and Consultants, Inc., April 1978.

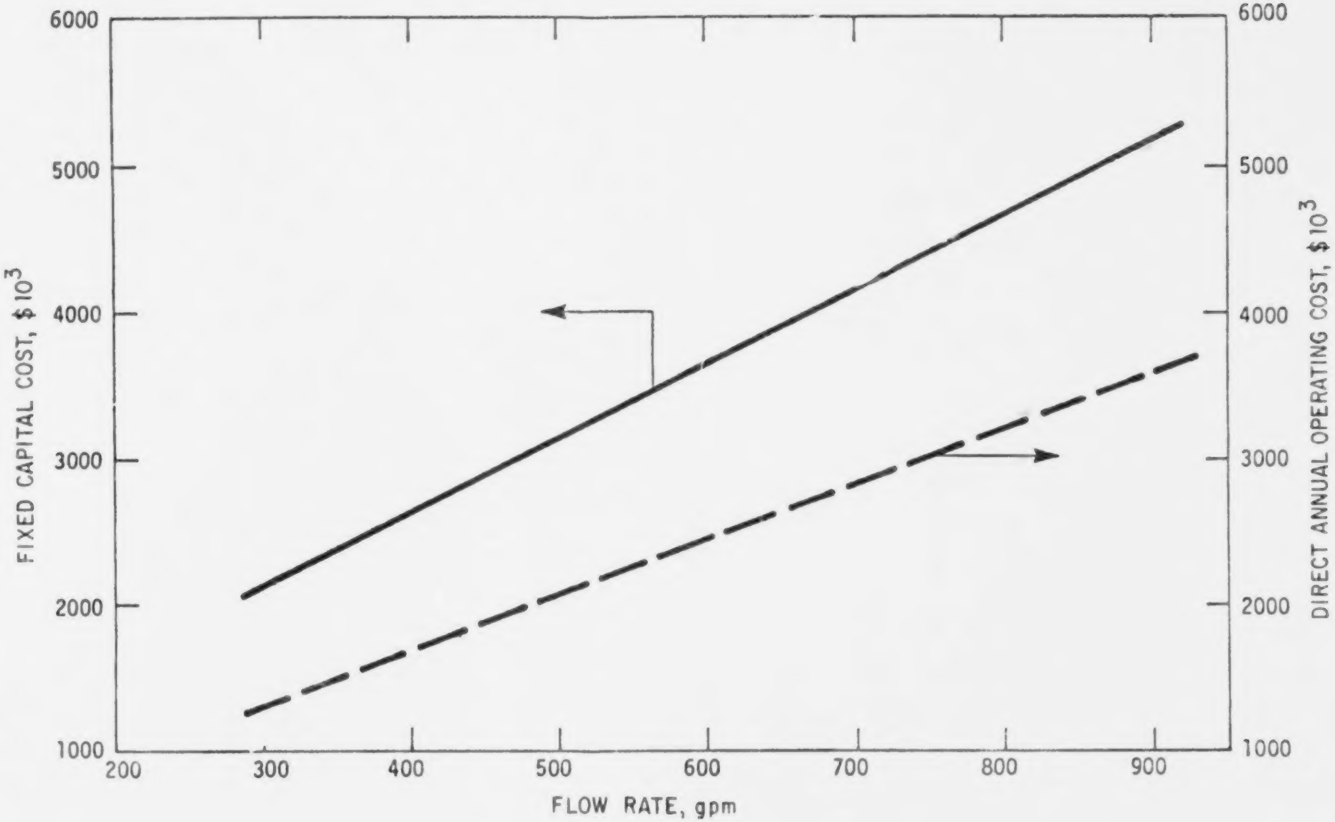
TABLE 5.2-9. COST OF AMMONIA RECOVERY

Item	Unit	Quantity
Fixed Capital Cost	$\$10^3$	
Towers		1,660
Heat exchangers		1,920
Drums, etc.		47
TOTAL		3,627
Direct Annual Operating Cost	$\$10^3$	
Maintenance @ 4%[a]		118
Labor, 24 hr/day @ $30/hr		237
Steam @ $3/MMBtu		1,565
Cooling water @ 3¢/m^3 circulated		60
Electricity @ 3¢/kW-hr		11
Chemicals		
NaOH		404
H_3PO_4		24
TOTAL		2,419
Credit for Ammonia Sales @ $110/ton	$\$10^3$/yr	816
Total Annual Control Cost[b]	$\$10^3$	2,395

[a] Maintenance is based on the fixed capital cost less contingency.

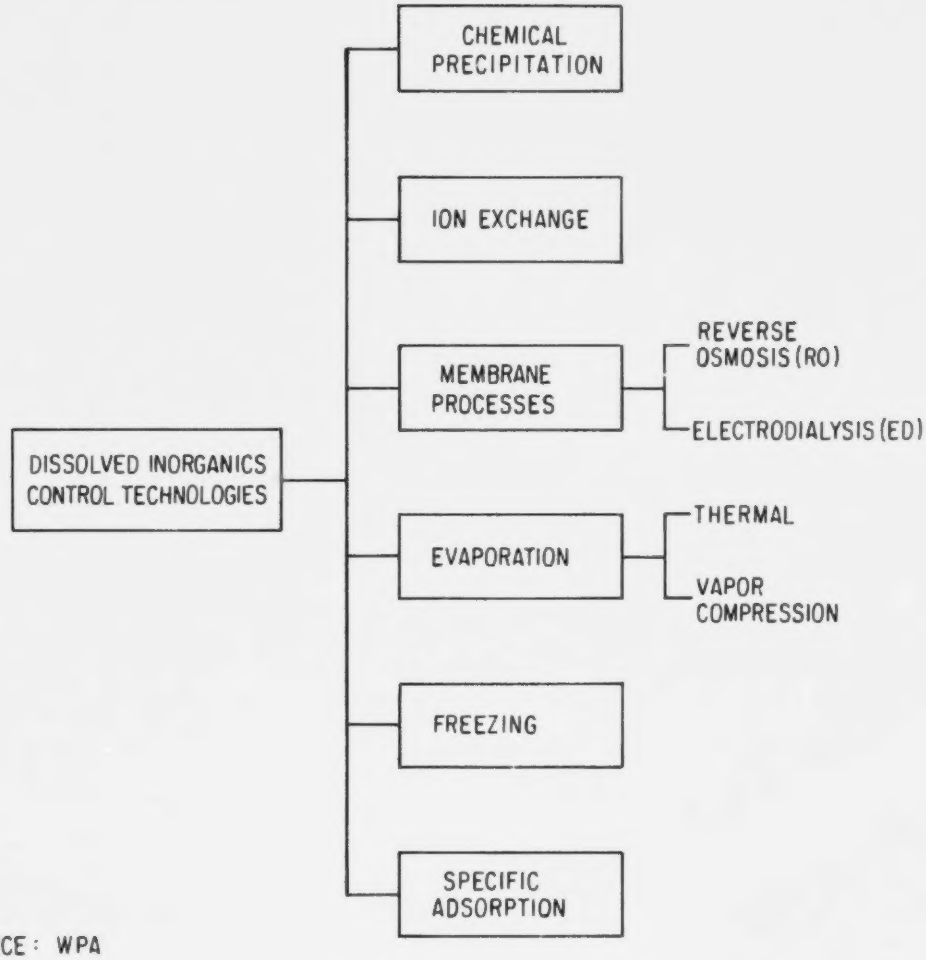
[b] See Section 6 for details on computation of the total annual control cost.

Source: WPA estimates based on information provided by U.S.S. Engineers and Consultants, Inc., April 1978.



SOURCE: WPA

FIGURE 5.2-7 COST OF AMMONIA RECOVERY WITH PHOSAM-W PROCESS


CHEMICAL PRECIPITATION
ION EXCHANGE
MEMBRANE PROCESSES
REVERSE OSMOSIS (RO)
ELECTRODIALYSIS (ED)
DISSOLVED INORGANICS CONTROL TECHNOLOGIES
EVAPORATION
THERMAL
VAPOR COMPRESSION
FREEZING
SPECIFIC ADSORPTION

SOURCE: WPA

FIGURE 5.2-8 DISSOLVED INORGANICS CONTROL TECHNOLOGIES

TABLE 5.2-10. KEY FEATURES OF CONTROL TECHNOLOGIES FOR DISSOLVED INORGANICS

Control Technology	Operating Principle	Components Removed	Removal Efficiency	Feed Requirements/ Restrictions	By-products and Wastes	Comments
Chemical Precipitation	Use of agents to promote the precipitation of inorganic solids from wastewaters.	Ca, Mg, heavy metals, alkalinity.	Variable, depending on constituents.	Lime, polymer, and soda ash may be required.	Sludge contaminated with heavy metals.	Generally followed by filtration and/or activated carbon adsorption.
Ion Exchange	Substitution of H^+ and OH^- ions for objectionable ionic species. Exchange resins regenerated with acid base or salt solutions.	Heavy metals, F^-, CN^-, scaling species.	90+% for most ions. Regeneration frequency is a key parameter.	Regenerants, replacement resins.	Spent regenerants and resins.	Most effective as a polishing process. Clearly applicable to boiler feedwater treatment needs; of limited use in treating process wastewaters containing high concentrations of organics or dissolved solids.
Membrane Processes (RO, ED)	Separation of dissolved matter by a semipermeable membrane under a pressure (RO) or electric (ED) gradient.	Ionized salts.	90-99% removal of dissolved salts.	Filtration, pH adjustment, foulant control.	Concentrate, spent membranes.	RO and ED have been used commercially for desalination. Concentrate stream may be 10-30% of input stream.
Evaporation (Thermal, Vapor Compression)	Application of heat (solar, steam, etc.) to evaporate wastewater or concentrate streams.	All nonvolatile species will remain in brine.	99+% rejection of nonvolatile dissolved solids.	Fouling/scaling of heat exchange surfaces must be prevented.	Recovered condensate, noncondensible gases, waste brine.	Solar evaporation may be unacceptable due to air pollution. Vapor compression evaporation has been successfully tested on retort waters.
Freezing	Cooling with formation of ice which is separated from remaining brine.	Dissolved salts, including organics.	90+% possible.		Concentrate stream.	Not yet demonstrated commercially.
Specific Adsorption	Adsorption of specific ions onto resins or other adsorbent.	Boron, fluoride, trace metals.	90+% in properly designed systems.	As above.	As above.	Useful as a polishing process.

Source: WPA.

Reverse osmosis (RO). Sometimes called "hyperfiltration," RO is a process for recovering relatively pure water from solutions. Water is passed through a hyperfilter, or semipermeable membrane, which rejects dissolved materials. As in normal filtration, the driving force is hydrostatic pressure, but in this case, the pressure has to be greater than the osmotic pressure of the solution. Osmotic pressures are related to the total molar concentration of the solution and its temperature (Hicks and Liang, January 1981).

The water is passed under pressure (greater than 200 psi) through a membrane which is impermeable to most inorganic salts and many organics. These "rejected" substances remain in a concentrate stream which may be 10-20% of the feedwater volume. The treated water or permeate will generally contain less than 10%, and often less than 1%, of the rejected substances. Costs scale primarily with the volume of water to be treated but are also dependent on concentration. At very high solute concentrations (e.g., seawater), costs increase rapidly due to the high applied pressures that are required. The flux of water through the membrane, i.e., the permeate recovery rate, increases linearly with the pressure by which the applied pressure exceeds the osmotic pressure. Fluxes of 10 gal/ft^2/day have been measured for retort water at an applied pressure of 600 psi. Typical applied pressures for brackish waters range from 200 to 600 psi and greater.

Membranes consist essentially of a thin skin (0.1 to 0.25 μm) of active chemical (cellulose acetate, polyamide) on a porous substructure, which may then be housed in a spiral-wound module for commercial application. Other geometries are also available. Rejection of strong electrolytes is normally in excess of 90% and can exceed 99 percent. Nearly complete rejection is obtained from most species with molecular weights greater than about 150. However, low molecular weight nonelectrolytes (e.g., small organic molecules like urea, and weak acids such as boric acid) are poorly rejected. Rejections of these substances can sometimes be improved by adjusting the solution pH to a value where the compound dissociates (e.g., boron is rejected above pH = 10).

Some advantages of RO treatment are the low labor and space requirements and the high rejection rates obtained for a wide range of dissolved contaminants. Of particular relevance to oil shale retort water is that both organic and inorganic compounds can be simultaneously removed under favorable pH conditions and that such a system can accommodate changing water flow rates. A serious disadvantage of the process is that the membranes are susceptible to blockage by deposition of solids. This so-called fouling results from solids present in the feed solution or from precipitation of solids as the concentration in the brine exceeds the solubility limit; it may even result from biological activity on the membrane surface.

Fouling rates may be reduced by proper pretreatment and by reducing the concentration increase in the brine. Reverse osmosis does not destroy the pollutants, it merely concentrates them into a smaller liquid stream. Reducing the concentration increase implies reducing the product recovery and increasing the amount of brine for disposal. Fouling can be further controlled by periodic washing, although there is generally a certain amount of irreversible fouling that determines membrane life and operating costs.

Costs scale proportionately with the volume of product water recovered, but they are also dependent on the degree of recovery and membrane fouling characteristics. As the concentration of pollutants in wastewater increases, so does the osmotic pressure; hence, higher applied pressures are required to maintain the desired permeate flux. Energy costs, however, are normally small relative to membrane costs.

Electrodialysis (ED). Electrodialysis is the use of an electromotive force to transport ionized materials in a solution through a diaphragm, or membrane. The process can be made selective by using ion-specific membranes which allow passage of only certain ions. A common application of electrodialysis is in the desalting of brackish waters containing 1,000-5,000 ppm of salts. A removal efficiency of 90-99% is usually achievable.

Thermal evaporation. This approach includes processes in which heat is applied to vaporize water, leaving a concentrated solution or slurry for disposal. The high energy required for evaporation is recovered in most processes by condensing the water vapor and, as a result, producing a stream of relatively pure water. Volatile contaminants, if present, may require removal in an upstream stripping process in cases where a clean product water is necessary. Multiple effect boiling (MEB) and multistage flash (MSF) are two procedures commonly used for evaporation (Water Purification Associates, December 1975).

Disadvantages of thermal processes are that volatile substances are not controlled, and (energy) costs are generally higher than for processes not involving a phase change. Problems related to scaling of heat transfer surfaces and corrosion are also encountered. These problems may be accentuated with waters containing high organic loadings, such as oil shale wastewater. Thermal processes may find application if there is a need for dirty steam, as occurs in many in situ processes.

Vapor compression evaporation. This is a method for evaporating water by the use of mechanical energy. Thermal energy required for evaporation is obtained by mechanical compression of the vapor instead of by heating. The wastewater is boiled in an evaporator to produce a vapor which is compressed in order to raise its temperature, and then it is passed through the tubes in the evaporator where the necessary heat exchange between the vapor and wastewater takes place. The vapor cools and condenses upon heat exchange and a relatively pure water is produced.

The advantage of vapor compression is that the heat required for vapor formation is recirculated so that the amount that must be dissipated is much less than the latent heat of vaporization. This approach results in relatively low energy requirements and essentially negligible cooling water requirements. The penalties are the high capital costs associated with the compressor, which must handle the large volumes of vapor, and increased maintenance costs. Other disadvantages of vapor compression evaporation are similar to those of the thermal processes.

The energy required for the single effect vapor compression units is about 70-90 kW-hr per thousand gallons of product water. Some single effect vapor compression units (RCC evaporator) can recover up to 98% of the wastewater containing up to 11,000 mg/l total dissolved solids.

Freezing. The water is reduced in temperature to produce a solid (ice) phase and a concentrated brine. The ice is washed free of salts and then melted to produce a virtually pure water. Both inorganics and organics are removed in the brine stream. Since the costs scale with the volume of water to be treated, freezing would normally be applied to relatively concentrated low volume wastes. While this process is theoretically more efficient than evaporation, it has yet to be applied commercially. It is included in this inventory as it may be useful for controlling retort waters, provided operating problems can be resolved in the future (Barduhn, September 1967; Water Purification Associates, December 1975).

Specific adsorption. The processes in this category are similar to the ion exchange processes, except that the affinity between the sorbent materials and the solutes being removed is of a physical nature. The sorbents may be natural or synthetic and usually have pores, or lattice vacancies, of uniform size and dimensions which are specific for the solutes. The processes are not applicable to high strength wastewaters and are generally used for trace removal applications.

Control Technologies Analyzed--

The following streams may require control of dissolved inorganics:

- Boiler Feedwater (stream 94)
- Cooling Tower Makeup Water (stream 97).

Based on the quality of the water, demineralization using reverse osmosis was examined as the most economical treatment of the mine water. A relatively large boiler blowdown is required, however, to maintain acceptable concentration levels in the boilers in order to prevent scaling. The boiler blowdown is used for processed shale moistening. The blowdown does represent an energy loss from the boiler system, and some heat recovery from this stream might prove cost effective. The material rejected by reverse osmosis is also used for processed shale moistening after equalization with other wastewaters. Table 5.2-11 gives the basis for design and costs of boiler feedwater treatment, and Figure 5.2-9 shows a specific cost curve for boiler feedwater treatment by reverse osmosis. This treatment could be considered as part of the process rather than pollution control.

Clarified mine water is used as cooling tower makeup. As a treatment, some sulfuric acid is added to convert calcium carbonate to the more soluble calcium sulfate. The cooling tower is operated at 1.5 cycles of concentration, which means that the concentration of dissolved species in the blowdown is 1.5 times that in the makeup. Since this concentration is not excessive, there should not be any problem in using the cooling tower blowdown for processed shale moisturizing. Table 5.2-12 contains design and cost information for the cooling tower makeup treatment, and Figure 5.2-10 presents a

TABLE 5.2-11. DESIGN AND COST OF BOILER FEEDWATER TREATMENT[a]

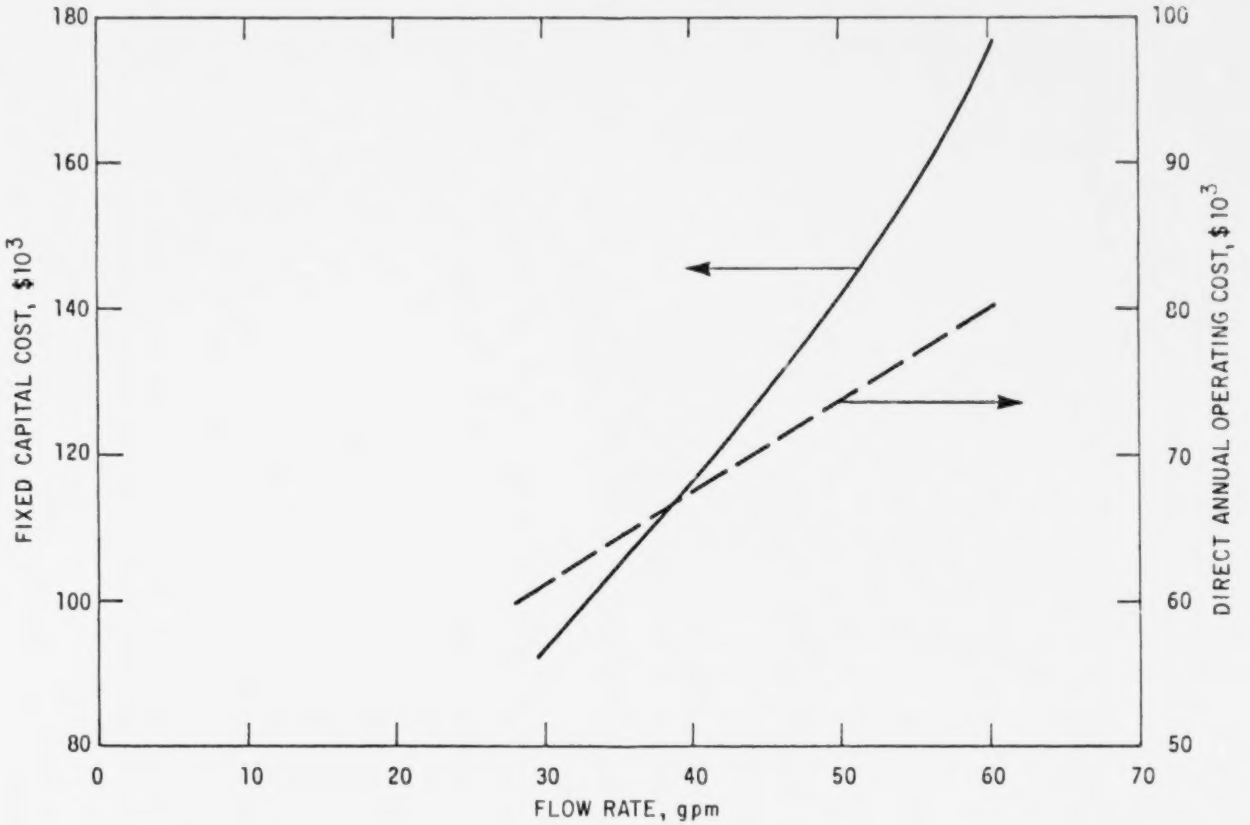
Item	Unit	Quantity
Boiler Blowdown	gpm	21
Steam Losses	gpm	11
Softener Regeneration Waste	gpm	11
TOTAL MAKEUP (clarified mine water)	gpm	43
Fixed Capital Cost	$\$10^3$	
Elements @ $1,160 each		20
Pressure vessel @ $1,920 each		6
Degasifier		5
Subtotal		31
Total equipment cost (250% of subtotal)		78
Civil work & installation (25% of total equipment cost)		19
Contingency		25
TOTAL		122
Direct Annual Operating Cost	$\$10^3$	
Maintenance @ 3%[b]		4
Labor, 4 hr/day @ $30/hr		40
Electricity @ 3¢/kW-hr		11
Membrane replacement (1.5-yr life) and chemicals		14
TOTAL		69
Total Annual Control Cost[c]	$\$10^3$	94

[a] This technology could be considered as part of the process rather than pollution control.

[b] Maintenance is based on the fixed capital cost less contingency.

[c] See Section 6 for details on computation of the total annual control cost.

Source: WPA estimates based on information from Peters and Timmerhaus, 1980.



SOURCE: WPA

FIGURE 5.2-9 COST OF BOILER FEEDWATER TREATMENT WITH REVERSE OSMOSIS

cost curve for the treatment. The cooling tower makeup treatment could be considered as part of the process rather than pollution control.

TABLE 5.2-12. DESIGN AND COST OF COOLING WATER TREATMENT[a]

Item	Unit	Quantity
Evaporation and Drift Losses	gpm	892
Blowdown	gpm	1,784
TOTAL MAKEUP (clarified mine water)	gpm	2,676
Cycles of Concentration	--	1.5
Sulfuric Acid Addition	mg/l (ppm) ton/yr	150 785
Direct Annual Operating Cost	$\$10^3$	
Sulfuric acid @ $65/ton		51
Total Annual Control Cost[b]		52

[a] This technology could be considered as part of the process rather than pollution control.

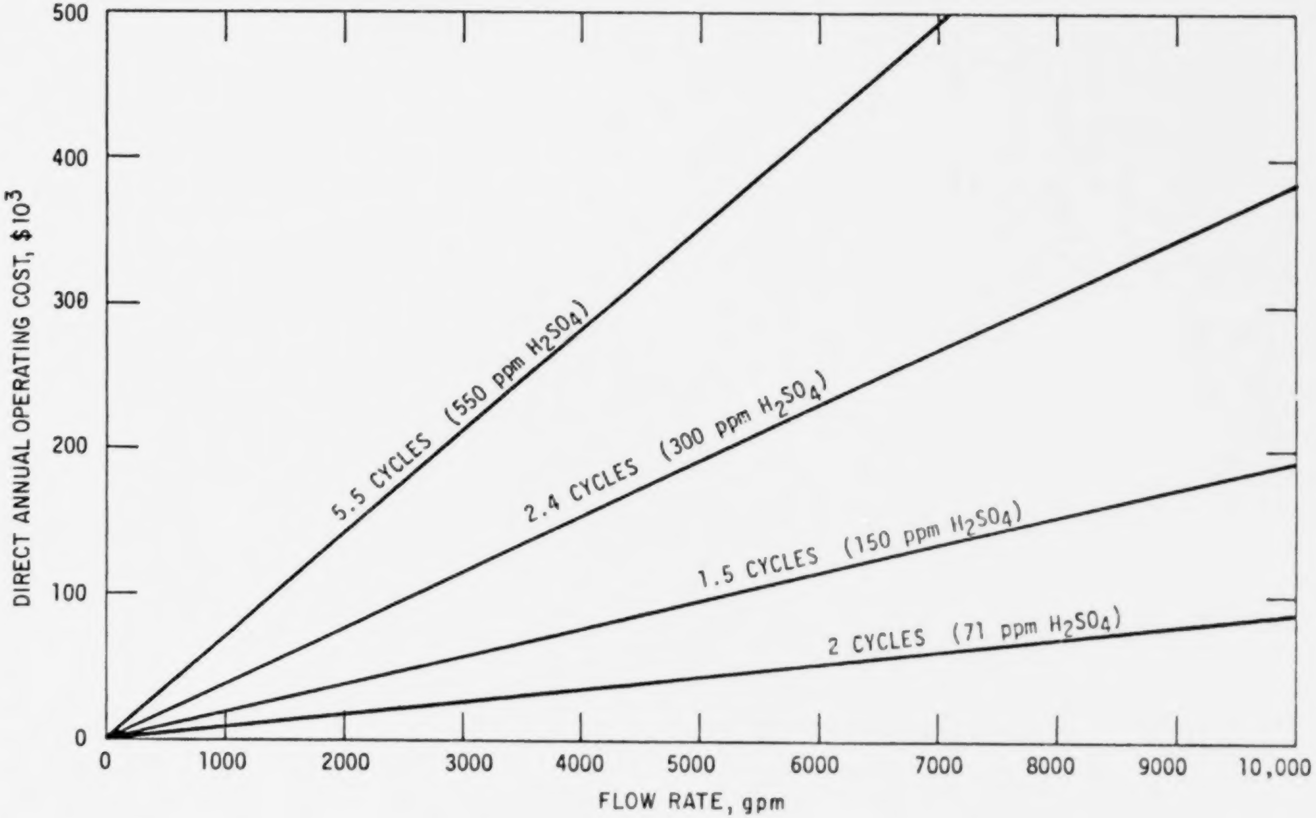
[b] See Section 6 for details on computation of the total annual control cost.

Source: WPA estimates based on information from Peters and Timmerhaus, 1980.

Other Control Technologies Analyzed--

Several additional dissolved inorganics control technologies were analyzed. These include reverse osmosis, boron adsorption, and phenol adsorption to remove dissolved salts, boron and phenol, respectively, from the excess mine water prior to its discharge. Cooling towers and solar evaporation ponds were examined for treating the process waters. Although these technologies have not been proposed for the Lurgi-Open Pit plant, they were analyzed as viable alternatives in the event that the wastewater disposal and reuse strategies for the plant are varied.

As stated earlier, the approach adopted for excess mine water disposal is to discharge it on the surface. If the quality of the excess mine water after clarification does not satisfy the criteria for surface discharge, the gross inorganic content can be reduced first by reverse osmosis (RO), followed by the removal of boron and phenol from the RO permeate using specific ion exchange resins.



SOURCE: WPA

FIGURE 5.2-10 COST OF COOLING WATER TREATMENT

Reverse osmosis is a useful technology in that it affords simultaneous removal of the dissolved inorganics and organics. With this technology, the wastewater is forced through a semipermeable membrane which allows the water to pass through but rejects the dissolved matter, especially that which is highly ionized. At optimum pH, up to 95% of the inorganics and organics can be rejected. The permeate is usually a fairly clean water that is suitable for high quality water needs. The RO technology has been tested on the aquifer waters from Tract C-b and a rejection of over 98% of the total dissolved solids has been obtained (Water Purification Associates, unpublished). The resin adsorption technologies are widely used in wastewater treatment, although experience with the aquifer waters from Tract C-a has not been documented. Two flow schemes (Examples I and II) depicting the above treatment and water reuse technologies are presented in Figure 5.2-11, while the flow diagrams for the RO process and the boron and phenol adsorption processes are presented in Figures 5.2-12 and 5.2-13, respectively. Table 5.2-13 gives the mine water composition before and after these treatments. Design and cost information for the RO process is presented in Table 5.2-14 and for the boron and phenol adsorption systems in Tables 5.2-15 and 5.2-16, respectively. The cost curves for the three technologies are illustrated in Figures 5.2-14, 5.2-15 and 5.2-16.

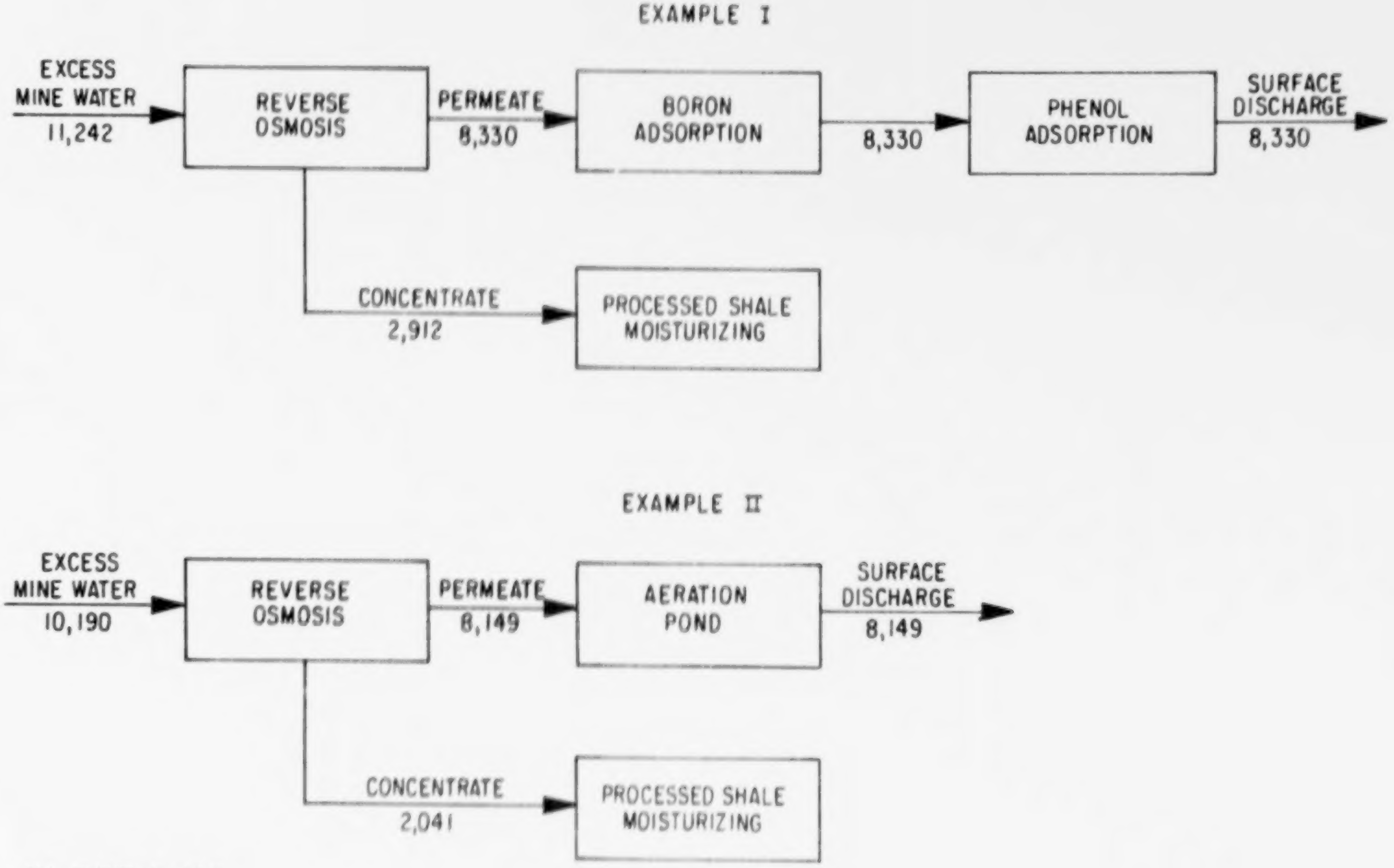
In the event that the process generated waters are not used for processed shale moisturizing, then a water reuse plan would have to be developed. One approach among many possibilities would be to treat the gas liquor (after ammonia removal) by adsorption on activated carbon to reduce the organic content. The treated water could then be used as cooling tower makeup water, thereby controlling the dissolved inorganics. Since the cooling tower can be run at fairly high cycles of concentration, most of the water is lost as evaporation and drift, and a small amount of blowdown is produced. The blowdown could then be placed in a solar evaporation pond to evaporate the remainder of the water, and the precipitated material could be properly discarded. Figure 5.2-17 shows this train for the gas liquor treatment. Table 5.2-17 presents the material balance around the cooling tower, while Tables 5.2-18 and 5.2-19 give the design and cost details for the cooling tower makeup treatment and solar evaporation pond, respectively. The cost curve presented previously in Figure 5.2-10 is applicable to the cooling tower makeup treatment indicated here. This treatment could be considered part of the process rather than pollution control. A cost curve for the solar pond is presented in Figure 5.2-18.

5.2.4 Dissolved Organics

Removal of volatile organics by stripping may be sufficient for reuse of process waters in processed shale moisturizing; however, nonvolatile organic components are not removable by stripping. Therefore, for higher quality uses, further treatment may be necessary. Some of the available approaches are discussed below.

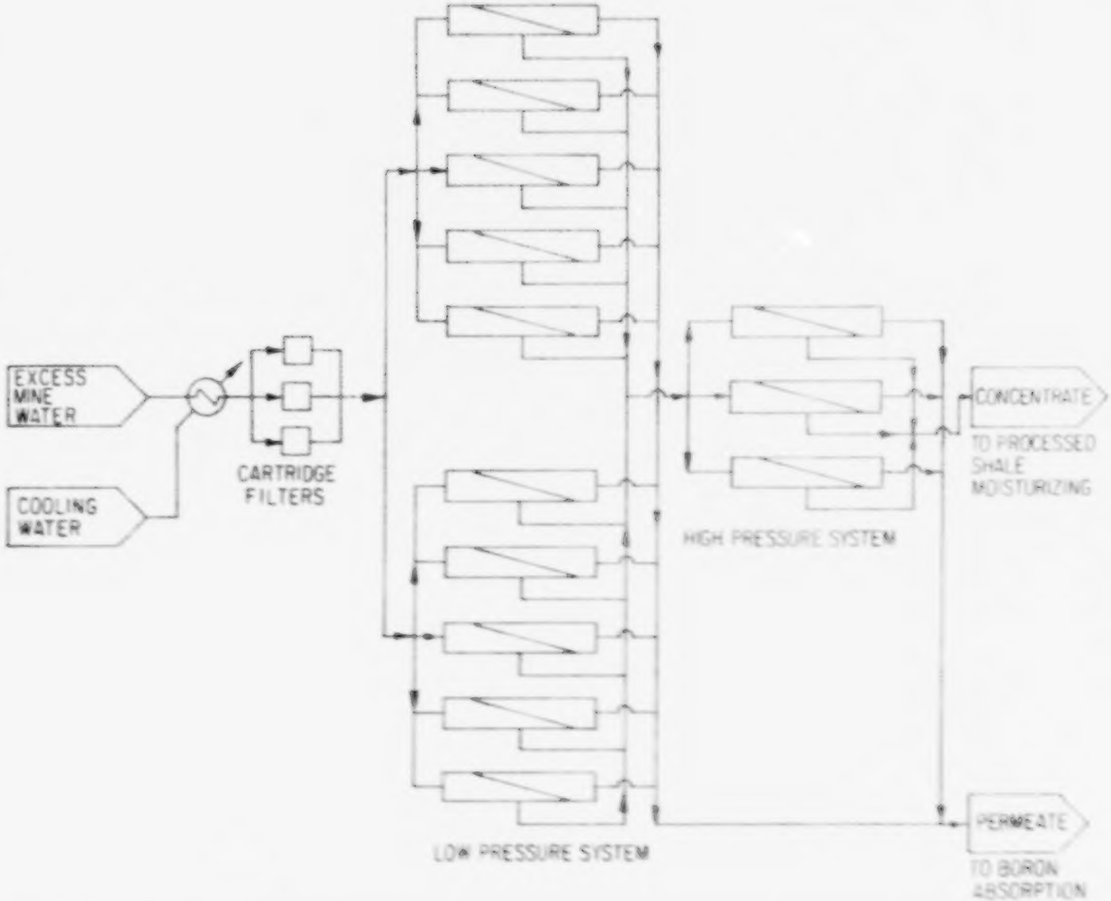
Inventory of Control Technologies--

The technologies available for dissolved organics control are shown in Figure 5.2-19 and are described in Table 5.2-20.


EXAMPLE I
EXCESS MINE WATER 11,242
REVERSE OSMOSIS
PERMEATE 8,330
BORON ADSORPTION
8,330
PHENOL ADSORPTION
SURFACE DISCHARGE 8,330
CONCENTRATE 2,912
PROCESSED SHALE MOISTURIZING
EXAMPLE II
EXCESS MINE WATER 10,190
REVERSE OSMOSIS
PERMEATE 8,149
AERATION POND
SURFACE DISCHARGE 8,149
CONCENTRATE 2,041
PROCESSED SHALE MOISTURIZING

ALL FLOWS IN GPM
SOURCE: WPA

FIGURE 5.2-11 FLOW SCHEME FOR RO, BORON ADSORPTION AND PHENOL ADSORPTION TREATMENTS

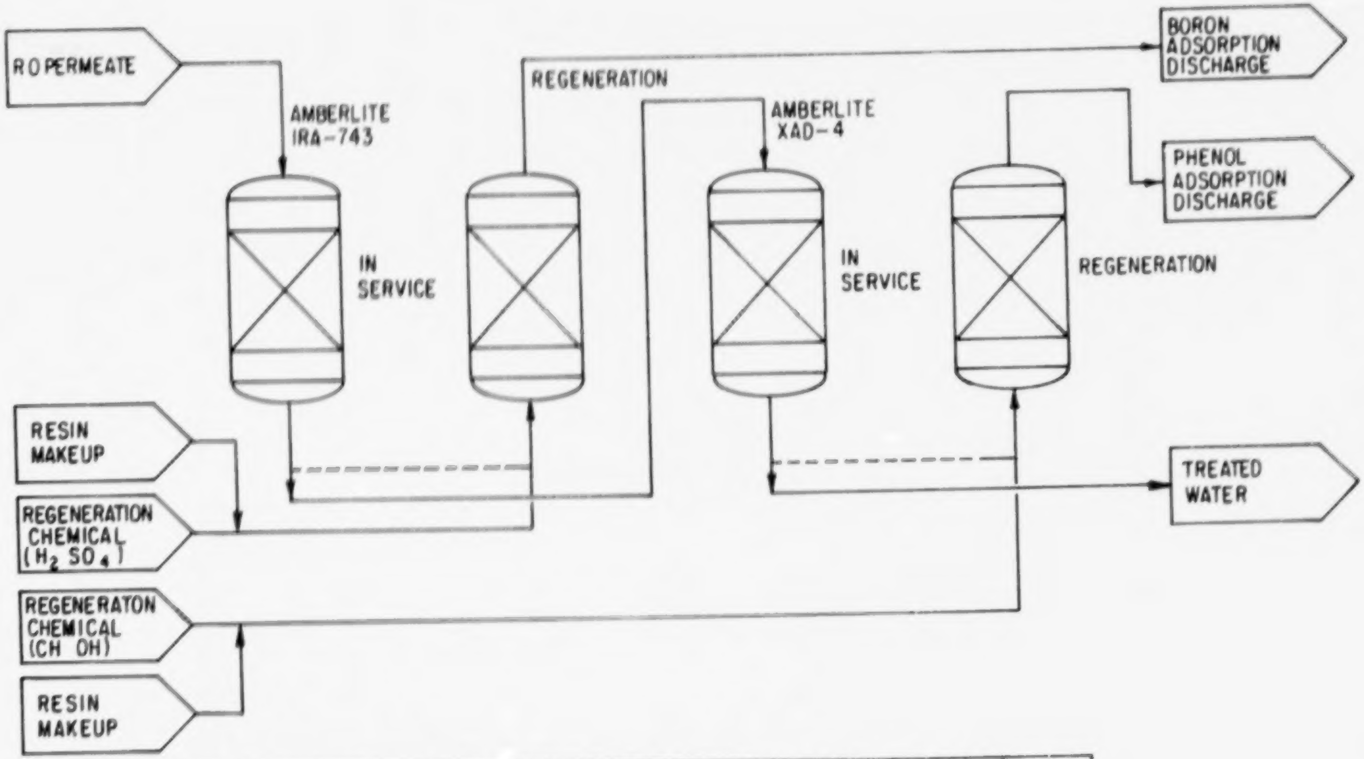


EXCESS MINE WATER
COOLING WATER
CARTRIDGE FILTERS
LOW PRESSURE SYSTEM
HIGH PRESSURE SYSTEM
CONCENTRATE
TO PROCESSED SHALE MOISTURIZING
PERMEATE
TO BORON ABSORPTION

STREAM IDENTITY	EXCESS MINE WATER	R O PERMEATE	R O CONCENTRATE	COOLING WATER
FLOW RATE: 10^3 lb/hr				
gpm	11242	8330	2912	N.D.
TEMPERATURE, °F	AMB	110	110	80
PRESSURE, psig	AMB	AMB	AMB	AMB

SOURCE: WPA

FIGURE 5.2-12 REVERSE OSMOSIS PROCESS FLOW SCHEME


RO PERMEATE
AMBERLITE IRA-743
IN SERVICE
REGENERATION
AMBERLITE XAD-4
IN SERVICE
REGENERATION
BORON ADSORPTION DISCHARGE
PHENOL ADSORPTION DISCHARGE
TREATED WATER
RESIN MAKEUP
REGENERATION CHEMICAL (H_2SO_4)
REGENERATON CHEMICAL (CH OH)
RESIN MAKEUP

STREAM IDENTITY	RO PERMEATE	RESIN MAKEUP	REGENERATION CHEMICAL (H_2SO_4)	REGENERATION CHEMICAL (CH_3OH)	RESIN MAKEUP	BORON ADSORPTION DISCHARGE	PHENOL ADSORPTION DISCHARGE	TREATED WATER
FLOW RATE:								
10^3 lb/hr		N.D.	0.03		N.D.	N.D.	N.D.	
gpm	8330			50 gpd				8330
TEMPERATURE, °F	110	AMB ←						→ AMB
PRESSURE, psig	AMB ←							→ AMB

SOURCE: WPA

FIGURE 5.2-13 BORON AND PHENOL ADSORPTION PROCESS FLOW SCHEME

TABLE 5.2-13. EXCESS MINE WATER COMPOSITION AFTER RO, BORON ADSORPTION AND PHENOL ADSORPTION TREATMENTS

Parameter	Raw Mine Water[a]	After Treatment, mg/l			
		RO Permeate	RO Concentrate	Boron Adsorption[b]	Phenol Adsorption[b]
Alkalinity, as $CaCO_3$	560	28	2,688	28	28
Aluminum	0.2	0.01	1.0	0.01	0.01
Ammonia, total	0.89	0.22	3.6	0.22	0.22
Arsenic	0.01	0.0005	0.05	0.0005	0.0005
Boron	0.62	0.31	1.9	~0	~0
Calcium	20	0.2	99.2	0.2	0.2
Chloride	18	0.9	86.4	0.9	0.9
Chromium	<0.01	<0.0005	0.05	<0.0005	<0.0005
COD	15	1.5	69	1.5	1.5
Cyanide	0.01	0.001	0.05	0.001	0.001
Fluoride	8.5	0.85	39.1	0.85	0.85
Lead	0.2	0.04	0.8	0.04	0.04
Mercury	0.003	0.0008	0.01	0.0008	0.0008
pH (units)	7.0	~7	~7	~7	~7
Phenols	0.0025	0.0013	0.01	0.0013	~0
Silica	20	4	84	4	4
Sodium	320	16	1,536	16	16
TDS	1,000	50	4,800	50	50
Sulfate	206	4.1	1,004	4.1	4.1
Sulfide	0.6	0.03	2.9	0.03	0.03
Flow Rate (gpm)					
Example I	(11,242)	(8,330)[d]	(2,912)	(8,330)	(8,330)
Example II[c]	(10,190)	(8,149)[d]	(2,041)	--	--

[a] Based on data in Table 4.2-22, assuming mine water is 43% from upper and 57% from lower aquifer.

[b] The removal efficiencies for very small concentrations of boron and phenol have not yet been established.

[c] In Example II, more of the mine water is used in processed shale moisturizing; therefore, a lower amount is available for treatment and disposal.

[d] Assuming permeate recovery factor is 80%.

Source: WPA estimates based on data from Gulf Oil Corp. and Standard Oil Co. (Indiana), May 1977.

TABLE 5.2-14. DESIGN AND COST OF REVERSE OSMOSIS TREATMENT OF EXCESS MINE WATER

Item	Unit	Example I	Example II[a]
Mine Water Flow	gpm	11,242	10,190
Number of Elements	--	5,000	4,530
Number of Pressure Vessels	--	800	730
Surface Area	ft^2/element	165	165
Membrane Flux	gal/day/ft^2	15-20	15-20
Electricity	kW	3,520	3,190
Fixed Capital Cost	$\$10^3$		
Elements @ $1,160 each		5,800	5,255
Pressure vessels @ $1,920 each		1,536	1,402
Subtotal		7,336	6,657
Total equipment cost (250% of subtotal)		18,340	16,643
Civil work and installation (25% of total equipment)		4,585	4,161
Contingency		5,275	4,796
TOTAL		28,200	25,600
Direct Annual Operating Cost	$\$10^3$		
Maintenance @ 4%[b]		917	832
Labor, 48 hr/day @ $30/hr		473	473
Electricity @ 3¢/kW-hr		832	754
Membrane replacement (1.5-yr life)		3,457	3,133
Scale inhibiting chemical		70	65
TOTAL		5,749	5,257

[a] In Example II, more of the mine water is used for processed shale moisturizing; therefore, a lower amount is available for treatment and disposal.

[b] Maintenance is based on the fixed capital cost less contingency.

Source: WPA estimates based on information from Hicks and Liang, January 1981.

TABLE 5.2-15. DESIGN AND COST OF BORON ADSORPTION SYSTEM

Item	Unit	Quantity
RO Permeate Flow Rate	gpm	8,330
Capacity of Resin (Amberlite IRA-743)	meq of boron/ml of resin	0.6
No. of Beds (2 standby)	--	9
Bed Volume	ft^3	3,000 (each)
Height of Column	ft	27
Flow Rate	gpm/ft^2	6
Sulfuric Acid Consumption[a]	ton/yr	127
Fixed Capital Cost	$\$10^3$	8,843
Direct Annual Operating Cost	$\$10^3$	
Maintenance @ 4%[b]		354
Resin replacement (5-yr life)		662
Labor, 24 hr/day @ $30/hr		237
Regeneration chemicals		8
Electricity @ 3¢/kW-hr		113
TOTAL		1,374

[a] The resin regeneration takes place every 50 days.

[b] Maintenance is based on the fixed capital cost less contingency.

Source: WPA estimates.

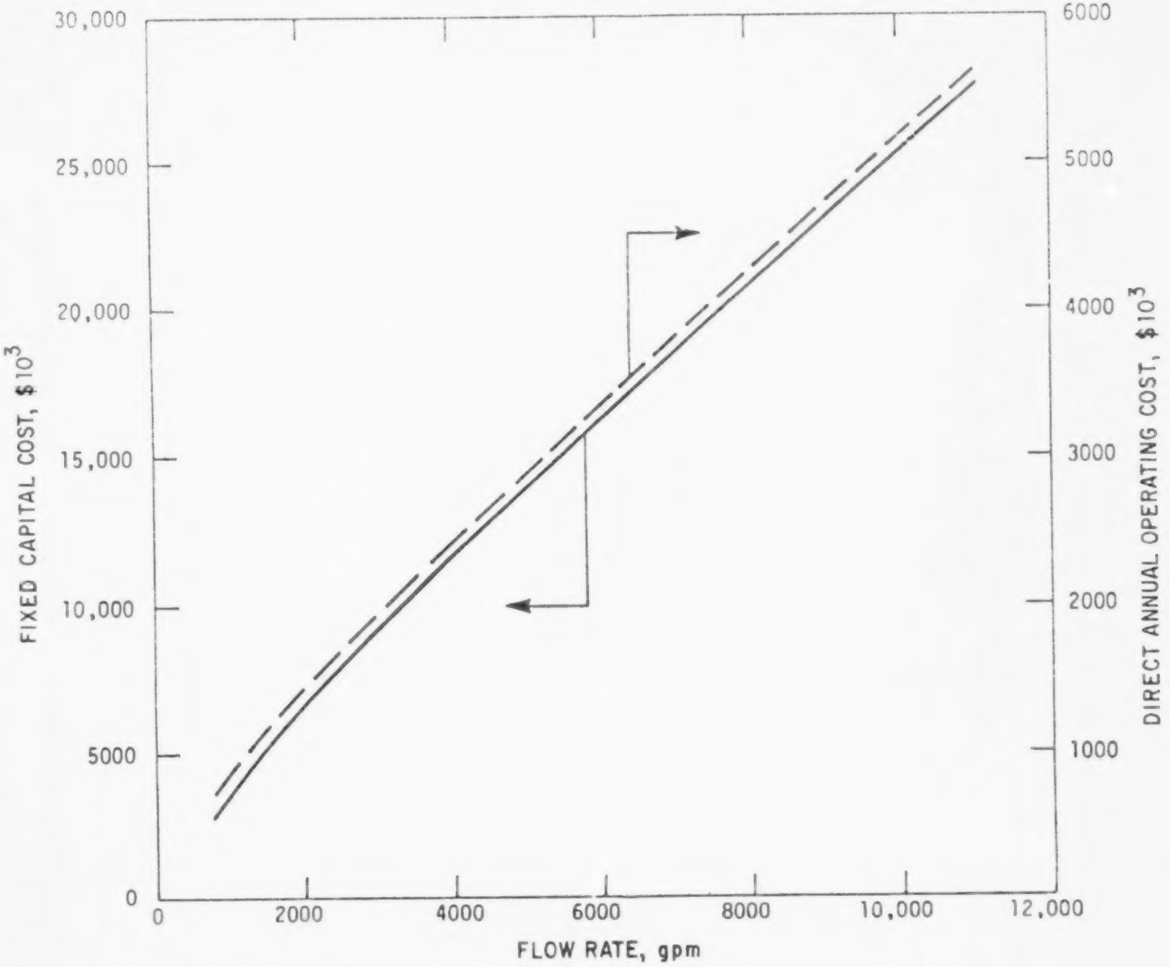
TABLE 5.2-16. DESIGN AND COST OF PHENOL ADSORPTION SYSTEM

Item	Unit	Quantity
RO Permeate Flow Rate	gpm	8,330
Type of Resin	--	Amberlite XAD-4
Capacity of Resin	lb phenol/ft^3 resin	0.78
Bed Volume	ft^3	3,000 (each)
No. of Beds (1 standby)	--	4
Cross Section Area	ft^2	200 (each)
Height of Column	ft	27
Service Rate	gpm/ft^3	1
Backwash Rate	gpm/ft^2	0.5
Methanol Consumption[a]	gal/day	50
Electricity Requirement	kW	311
Fixed Capital Cost	\$10^3	6,060
Direct Annual Operating Cost	\$10^3	
Maintenance @ 4%[b]		245
Resin replacement (5-yr life)		540
Labor, 24 hr/day @ \$30/hr		237
Regeneration chemicals		11
Utilities		74
TOTAL		1,107

[a] The resin regeneration takes place every 66 days.

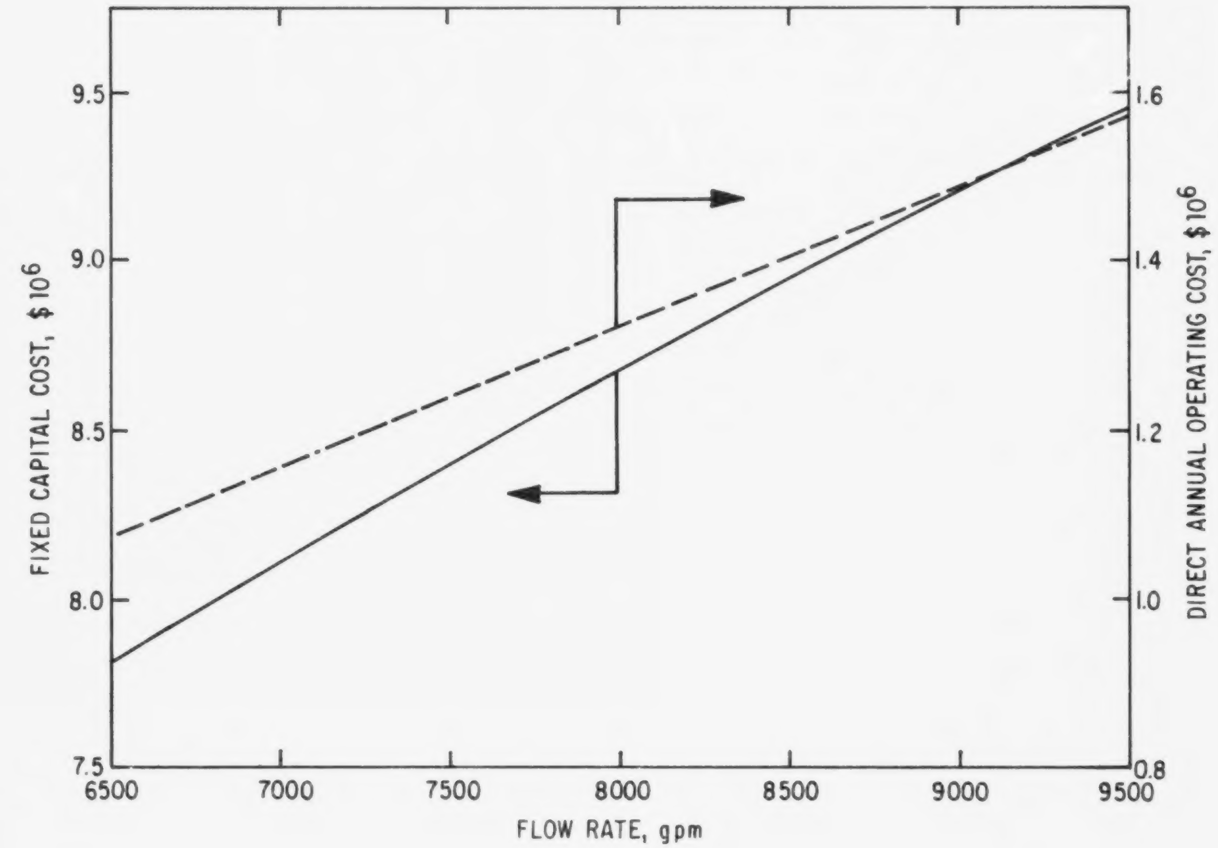
[b] Maintenance is based on the fixed capital cost less contingency.

Source: WPA estimates.



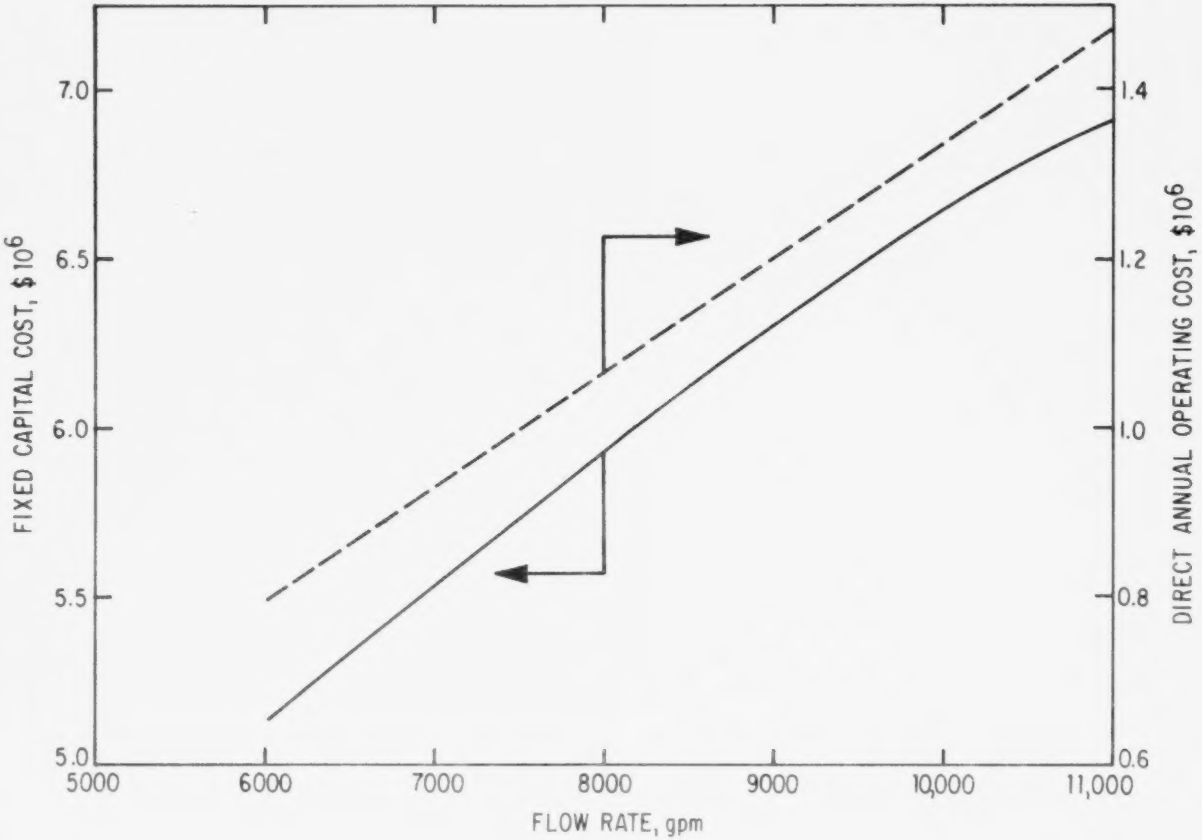
SOURCE: WPA

FIGURE 5.2-14 COST OF ORGANICS REMOVAL WITH REVERSE OSMOSIS



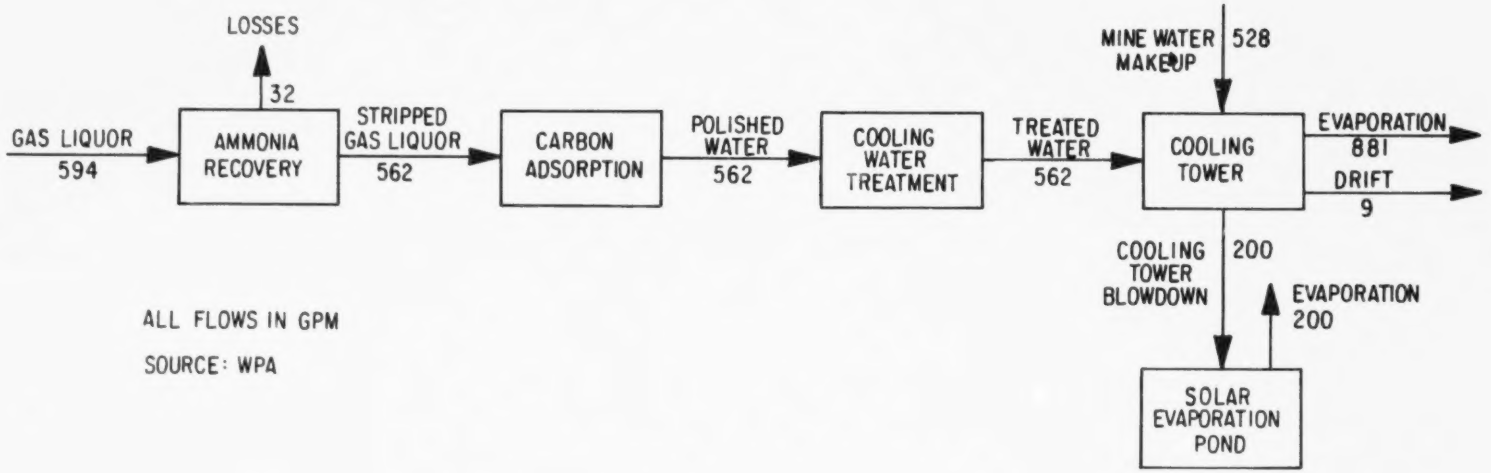
SOURCE: WPA

FIGURE 5.2-15 COST OF BORON REMOVAL WITH ION EXCHANGE SYSTEM



SOURCE: WPA

FIGURE 5.2-16 COST OF PHENOL REMOVAL WITH ION EXCHANGE SYSTEM


LOSSES
32
GAS LIQUOR
594
AMMONIA RECOVERY
STRIPPED GAS LIQUOR
562
CARBON ADSORPTION
POLISHED WATER
562
COOLING WATER TREATMENT
TREATED WATER
562
MINE WATER MAKEUP
528
COOLING TOWER
EVAPORATION
881
DRIFT
9
COOLING TOWER BLOWDOWN
200
EVAPORATION
200
SOLAR EVAPORATION POND

ALL FLOWS IN GPM

SOURCE: WPA

FIGURE 5.2-17 FLOW SCHEME FOR COOLING TOWER MAKEUP AND SOLAR EVAPORATION TREATMENTS

TABLE 5.2-17. MATERIAL BALANCE AROUND COOLING TOWER

	Before Treatment				After Treatment			
	Wastewater From Carbon Adsorption	Mine Water Makeup	Total		Evaporation	Drift	Blowdown to Solar Pond	
Components	lb/hr (gpm)	lb/hr (gpm)	Mass %	lb/hr (gpm)	lb/hr	lb/hr (gpm)	Mass %	lb/hr (gpm)
NH_3	6	--	0.001	6	6	--	--	--
TDS	429	265	0.127	694	--	--	0.688	694
Organics	85	--	0.016	85	--	--	0.084	85
H_2O	281,049 (562)	264,000 (528)	99.856	545,049 (1,090)	440,500	4,500 (9)	99.227	100,049 (200)
TOTAL	281,569	264,265	100.000	545,834	440,506	4,500	100.00	100,828

Source: WPA estimates.

TABLE 5.2-18. DESIGN AND COST OF COOLING TOWER MAKEUP TREATMENT*

Item	Unit	Quantity
Evaporation and Drift Losses	gpm	890
Blowdown	gpm	200
TOTAL MAKEUP	gpm	1,090
Cycles of Concentration	--	5.5
Sulfuric Acid Addition	mg/l (ppm) ton/yr	550 1,185
Direct Annual Operating Cost	$10^3	
Sulfuric acid @ $65/ton		77

* This technology could be considered as part of the process rather than pollution control.

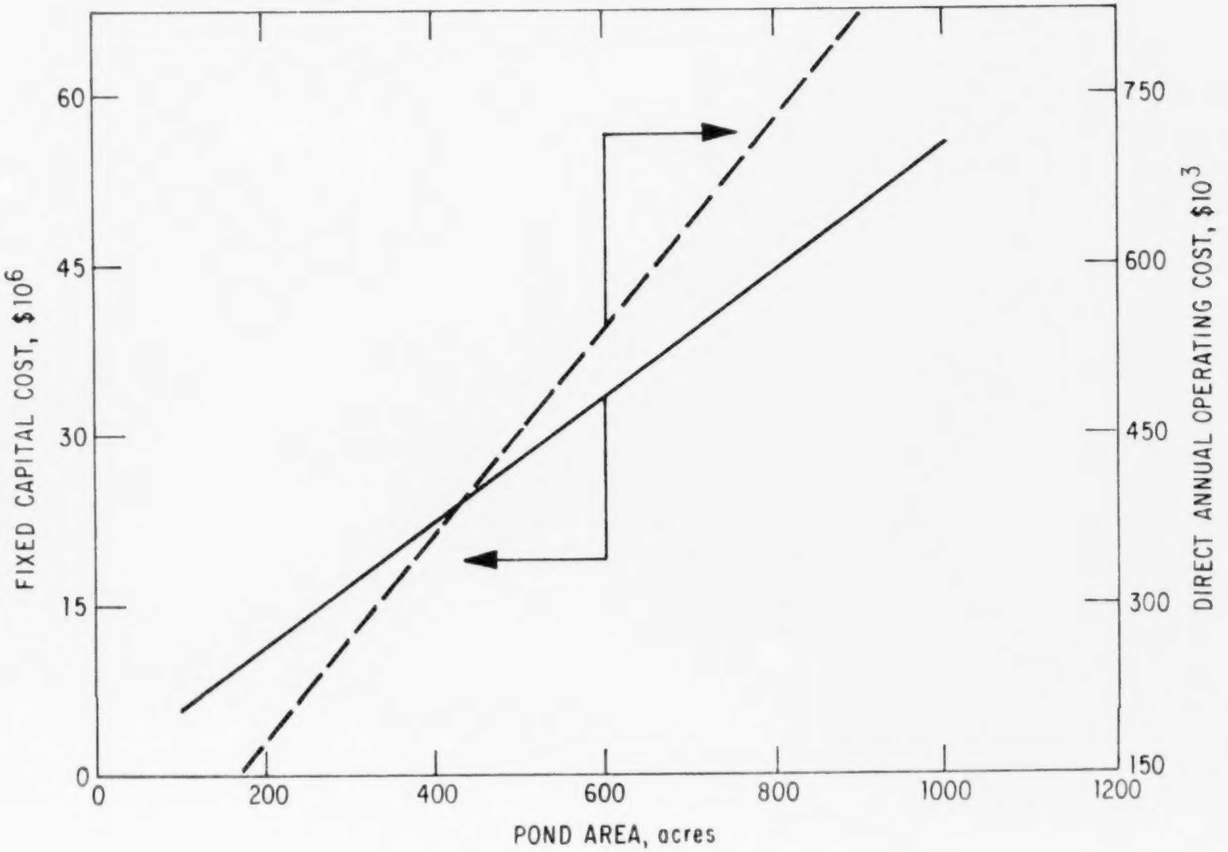
Source: WPA estimates based on information from Peters and Timmerhaus, 1980.

TABLE 5.2-19. DESIGN AND COST OF SOLAR EVAPORATION POND

Item	Unit	Quantity
Flow Rate to Pond	gpm acre-ft/yr	200 290
Evaporation Rate	in/yr	15
Pond Area	acres	257
Pond Depth	ft	3
Liner (chlorosulfonated polyethylene)	10^3 ft^2	11,200
Fixed Capital Cost	$10^3	14,200
Direct Annual Operating Cost	$10^3	
Maintenance @ 2%*		231

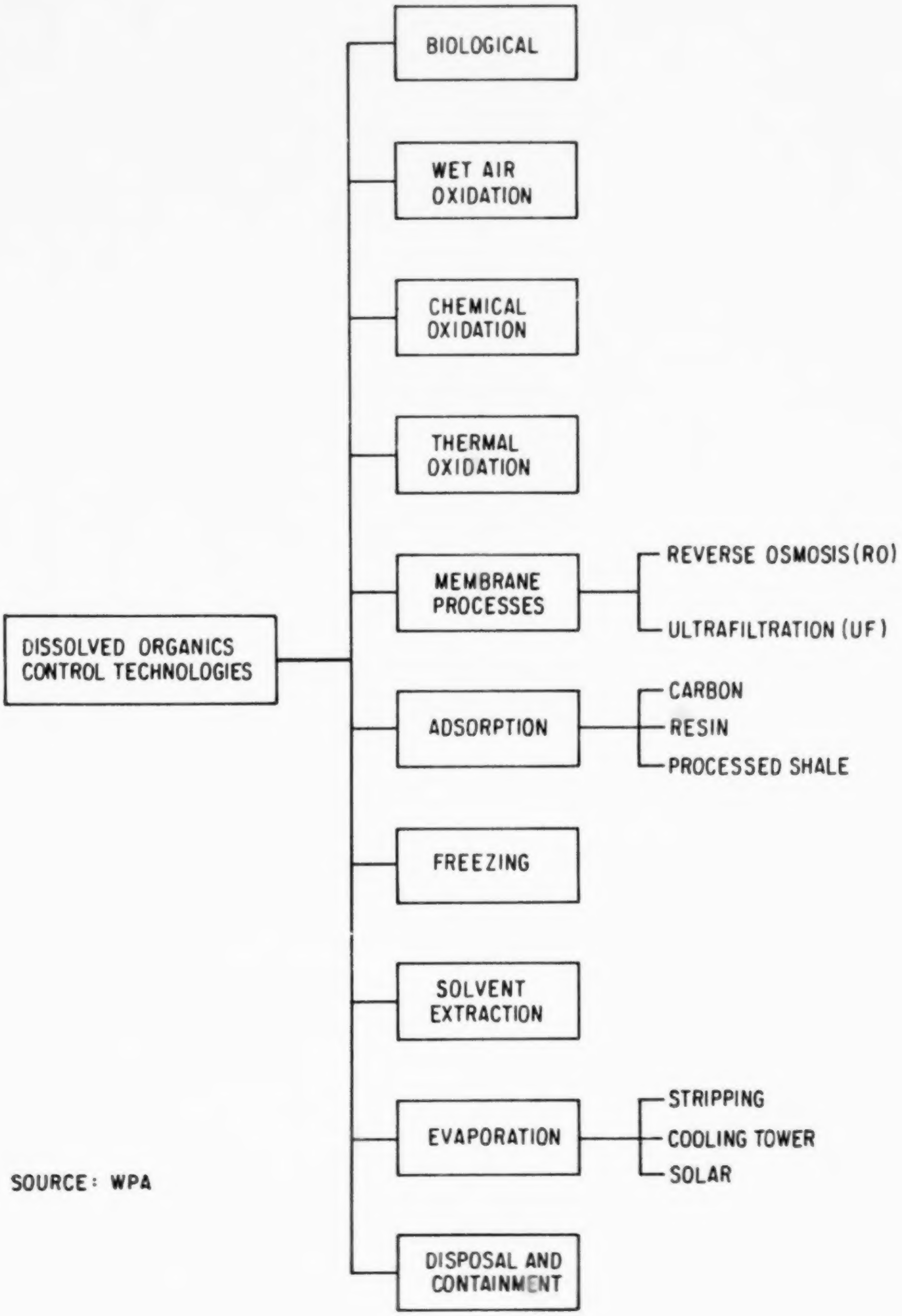
* Maintenance is based on the fixed capital cost less contingency.

Source: WPA estimates.



SOURCE: WPA

FIGURE 5.2-18 COST OF SOLAR POND


BIOLOGICAL
WET AIR OXIDATION
CHEMICAL OXIDATION
THERMAL OXIDATION
MEMBRANE PROCESSES
REVERSE OSMOSIS (RO)
ULTRAFILTRATION (UF)
DISSOLVED ORGANICS CONTROL TECHNOLOGIES
ADSORPTION
CARBON
RESIN
PROCESSED SHALE
FREEZING
SOLVENT EXTRACTION
EVAPORATION
STRIPPING
COOLING TOWER
SOLAR
DISPOSAL AND CONTAINMENT

SOURCE: WPA

FIGURE 5.2-19 DISSOLVED ORGANICS CONTROL TECHNOLOGIES

TABLE 5.2-20. KEY FEATURES OF CONTROL TECHNOLOGIES FOR DISSOLVED ORGANICS

Control Technology	Operating Principle	Components Removed	Removal Efficiency	Feed Requirements/ Restrictions	By-products and Wastes	Comments
Biological	Oxidation to CO_2 and H_2O (aerobic) or reduction to CH_4 (anaerobic) in the presence of suspended bacteria.	TOC, BOD, COD.	50% removal of TOC typical for retort waters. Efficiency enhanced by addition of PAC.	Relatively constant feed temperature and pollutant loadings are required to minimize "shocks" to the system. Air or oxygen must be added to aerobic systems. Supplemental nutrients may be required.	Biosludge, CO_2 in aerobic, CH_4 in anaerobic process.	Long residence times (days) require large reactor vessels. Air emissions during aeration may require that the vessels be enclosed.
Wet Air Oxidation	Direct reaction of O_2 with wastewater in a closed, pressurized vessel at elevated temperatures.	TOC, BOD, COD, as well as some oxidizable inorganics.	90+% removal of BOD, COD, TOC is possible in a system with a residence time of one hour or greater.	Air or oxygen; heat if autothermic reaction conditions are not present.	Vent gases containing CO, CO_2, light hydrocarbons, NH_3, sulfur species.	Promising, but not proven in this application. Fairly rigorous construction materials are required.
Chemical Oxidation	Reaction of organics in wastewater with O_3, peroxides or chlorine-based oxidants.	TOC, BOD, COD, oxidizable inorganics.	90+% achievable depending upon conditions of operation.	Oxidant.	Vent gases, wastewater and reaction products.	Chlorine-based oxidants may cause problems with treated wastewater.
Thermal Oxidation	Organics are combusted and the water stream is simultaneously evaporated.	All oxidizable organics.	Essentially 100% in properly designed system.	Feed should be concentrated to reduce fuel required for water evaporation.	Flue gases.	If NH_3 or sulfur species are present, NOx and SO_2 emissions may require control. Effective but expensive control.
Membrane Processes (UF, RO)	Separation of water and dissolved matter by semi-permeable membrane under influence of pressure field.	Large molecules (UF); intermediate size and ionizable molecules (RO).	50-98% of separable components.	Filtration, pH adjustment, removal of foulants.	Concentrate stream, spent membranes.	Long-term membrane fouling not yet studied.
Adsorption (Carbon, Resin, Processed Shale)	Adsorption of organics in water by activated carbon or polymeric resin. Powdered activated carbon has been used in conjunction with biological processes.	Many organics.	50% removal of TOC typical for raw and pretreated oil shale wastewaters.	Adsorbent.	Spent adsorbent.	Probably more effective as a polishing rather than a bulk organics removal process.

(Continued)

TABLE 5.2-20 (cont.)

Control Technology	Operating Principle	Components Removed	Removal Efficiency	Feed Requirements/ Restrictions	By-products and Wastes	Comments
Freezing	Cooling to form pure ice crystals which are separated from the concentrated brine.	TOC, TDS.	90+% possible.		Concentrate stream, ice.	Volatile components are removed along with the nonvolatiles. Not yet demonstrated commercially.
Solvent Extraction	Wastewater is intimately mixed with a water-immiscible organic solvent. Dissolved organics partition occurs between water and organic solvent phase.	Components soluble in organic solvent used.	Found to be ineffective for oil shale wastewaters.	Solvent, solvent regeneration system.	Recovered organics.	Will not be used unless suitable solvent is found.
Evaporation (Stripping, Cooling Tower, Solar)	Evaporate volatile components by applying heat via steam, solar energy, or exchange with the cooling water return from the plant. Simultaneously concentrate the nonvolatile compounds.	TOC, TDS.	Variable, depending on the volatility of the compounds.	Removal of volatile components preferred.	Overhead vapors and concentrate stream.	Direct steam stripping may remove azeotropic components. Slow air and biological oxidation are possible with the cooling tower and solar evaporation.
Disposal and Containment	Fixing of the contaminants on a substrate or disposal or containment with isolation from surroundings.	TOC, TDS.	Variable, depending upon the method used and surrounding factors.	Removal of volatile components preferred.		The wastewater may be contained, or reinjected, underground. Contaminants may be chemically and physically fixed on the processed shale.

Source: WPA.

Biological treatment. Biological processes may be aerobic, where organics are oxidized to carbon dioxide and water, or anaerobic, where the organics are reduced to methane. Both approaches produce sludge as a waste. Aerobic processes are faster and less susceptible to toxicity problems than anaerobic processes, but oxygenation equipment is required. Bench-scale tests on retort waters have shown that minor changes in retort water composition can result in a significant reduction in the performance of a well-acclimated system. In the presence of biorefractory (nonbiodegradable) organics, powdered-activated carbon may be added to the bioreactors to achieve acceptable reduction in organic content. Necessary pretreatment includes stripping, pH adjustment, and nutrient addition; control of specific toxic materials may be required as well (Adams and Eckenfelder, 1974; Hicks, et al., June 1979; Hicks and Wei, December 1980).

Wet air oxidation (WAO). This is a procedure for the destruction of organic matter dissolved or suspended in water or wastewater by oxidizing with air at high temperatures. The temperatures used are above the normal boiling point of water, and the reaction is carried out under pressure to prevent boiling. The pressure is usually 600 psig or above. The degree of oxidation achieved depends on the temperature and the material oxidized.

The advantage of WAO is that the organics do not have to be biodegradable to be oxidized. In fact, WAO often produces biodegradable substances from refractory material. For economic reasons, it is recommended that WAO systems be designed to remove no more than 80% of the organics. The optimum effluent is one that has a COD/BOD ratio of unity, i.e., the chemically oxidizable material is also biologically oxidizable. Biological oxidation can be used as a post treatment (Water Purification Associates, December 1975; Wilhelmi and Knopp, August 1979).

The WAO procedure is normally used for high strength wastes because costs scale with the volume of water to be treated. The energy needs for WAO often can be supplied by heat released in the process itself if the wastewater has a high concentration of reactive material. It is an expensive process and would be considered only for high strength wastes not amenable to other treatments, such as solvent extraction.

Chemical oxidation. In this process, oxidation of the organics is caused by adding oxidizing agents to the wastewaters. The oxidants are usually comprised of ozone, peroxides, chlorine, chlorates, etc. These chemicals are nonselective; that is, they oxidize total organic carbon as well as some inorganics. The oxidation may be carried out at ambient temperature, which is an advantage. Formation of obnoxious wastes is likely with chlorinated oxidants. Explosion is also a possibility under uncontrolled conditions.

Thermal oxidation. The wastewater is evaporated and the dissolved organics are simultaneously combusted by directly firing burners that are submerged under the wastewater. Organic nitrogen and sulfur compounds will convert to NOx and SO_2, which is a disadvantage. Additional waste gases may form if the fuel combustion is incomplete. Heat transfer within the wastewater is efficient; however, due to the presence of a large amount of noncondensable combustion gases, waste heat recovery from the overhead vapors may not be practical. Energy requirements can be reduced by using a preconcentrated wastewater.

Reverse osmosis. In addition to removing inorganics, this process removes organics to a certain extent, particularly if the organics are ionized. Tests on in situ retort waters have shown that, at a high pH, about 95% of the organics are removed. Modern polyamide thin film membranes are available for high pH operation, but additional data on membrane fouling characteristics with retort waters are required. The concentrate stream produced requires treatment, possibly by WAO (Water Purification Associates, December 1975; Hicks and Liang, January 1981).

Ultrafiltration. In addition to separation of oils and suspended particles, ultrafiltration will also separate large organic molecules (MWt ≧ 1,000). It is unlikely that ultrafiltration will be incorporated into a treatment train for the removal of large organic molecules, as these are not a significant fraction of total organics in retort waters. However, ultrafiltration may be used for emulsified oil separation and, in that case, would serve as a useful pretreatment to RO (Water Purification Associates, December 1975).

Carbon adsorption. This technology is used to remove organic materials from sewage and industrial water, as well as taste and odor from drinking water. It is usually used in conjunction with biological treatment as a pretreatment or polishing treatment (Cheremisinoff and Ellerbusch, 1978; Water Purification Associates, December 1975). Laboratory results from combined carbon adsorption and biological treatment of modified in situ oil shale retort water indicate that up to 85% removal of dissolved organics can be achieved compared to approximately 50% removal with biological treatment alone (Jones, Sakaji and Daughton, August 1982).

Activated carbon is produced by charring wood or coal at high temperatures. Charring temperature is the main factor determining the adsorption characteristics of granular or powdered-activated carbon.

Carbon must be regenerated when it is exhausted. The regeneration is accomplished by passing the carbon through a furnace at high temperature, usually around 800-1,000°C, with restricted oxidation to remove the adsorbed layer on the carbon. The quality of carbon after regeneration is slightly lower than the virgin carbon, and small quantities of virgin carbon must be added to retain the required activity.

Activated carbon has ion exchange groups and can be used to remove metal ions from water. It has been found that, under proper conditions of pH and oxidation, some metal ions are adsorbed very strongly.

Regeneration costs are a significant part of overall treatment costs, making the process uneconomical for high strength wastes, for which frequent regeneration is required. Regeneration also is not attractive for small units. Energy costs for running an activated carbon wastewater treatment plant are small, not considering regeneration, and are proportional to the pressure drop across the activated carbon contactor. Fouling in carbon adsorption units is reduced if the influent stream is adequately pretreated.

Resin adsorption. Resin adsorption is a physical process for removal of organic materials. Normally, it is considered as a polishing step, after bulk organic removal in upstream wastewater treatment steps, but may be used on waters having higher loadings than would be used for carbon. Also, it is useful for removal of specific toxic materials and phenol.

The polymer (resin) surface can be made hydrophobic or hydrophilic. Activated groups can be introduced to increase selectivity. Regeneration can be accomplished by washing with methanol, weak acid or weak base. Steam can be used to vaporize adsorbed materials.

Adsorption on processed shale. This method has been proposed for organics control in retort waters at oil shale plants. In studies at Lawrence Berkeley Laboratory, processed shale from the Lurgi, Paraho, TOSCO II, and three simulated in situ processes were contacted with four separate simulated in situ retort waters in batch and continuous (column) systems (Fox, Jackson and Sakaji, 1980). These studies indicated that the processed shale reduces the inorganic carbon by 50-98%, the organic carbon by 7-73%, and elevates the pH from initial levels of 8-9 to a final level of 10-11. An advantage of the process is that the increase in pH would facilitate downstream ammonia stripping and would reduce the loading on downstream organic removal steps.

Freezing. As previously discussed, freezing also removes dissolved organics. One advantage of freezing over evaporation processes is that volatile organics are removed as well. This process has yet to be applied commercially (Barduhn, September 1967; Water Purification Associates, December 1975).

Solvent extraction. When wastewater is contacted with a sparingly soluble immiscible organic solvent, the dissolved organic contaminants partition themselves between the aqueous and organic phases according to their relative solubility in each. The organic phase is separated and the dissolved contaminants removed in a distillation step. Alternatively, the solvent and dissolved organics may be incinerated. Solvent extraction is most economical for high strength wastes because costs scale with the volume of water to be treated and are relatively independent of the amount of substances removed. Unfortunately, effective solvents for the wide range of organics present in retort water have not been found, and it appears unlikely that solvent extraction will be useful in retort water treatment (Hicks, et al., June 1979).

Stripping. Volatile organics are removed along with ammonia and the acid gases in a stripping column or other thermal evaporative process. The amount of organics removed depends essentially on their volatility relative to water. Organics in retort water are relatively nonvolatile and indications are that less than 20% will be removed in a column stripping 99% of the ammonia. Organics in gas condensates, such as the TOSCO II foul water, are significantly more volatile, and bench-scale tests have shown that up to 85% of the organics are removed along with the ammonia. The volatile organics may then be incinerated, along with the other stripped gases, or may be adsorbed from the gas stream prior to ammonia recovery (Hicks and Liang, January 1981).

Cooling tower. The cooling tower may be regarded as a water treatment system. As such, its main function is to concentrate the dissolved salts, which may then be removed at lower cost in a sidestream or blowdown treatment stage. When using process wastewaters as cooling tower makeup, upstream removal of ammonia and organics need not be as efficient (and therefore as expensive) as when the wastewater is discharged. It has been demonstrated that refinery phenolic wastewaters can be used in a cooling tower and that bio-oxidation of phenol will occur with very high efficiencies (Hart, June 11, 1973). The conditions necessary for successful bio-oxidation are low sulfide (below 2 ppm) and small variations in pH (between 7.8 to 8.3). Chlorination is used to prevent biological growth. Corrosion of steel has been low. Ammonia will not concentrate in a cooling tower, but it will vaporize with the water.

Solar evaporation. Solar radiation incident upon the surface of an open evaporation pond is used as the energy source. Large, lined, shallow ponds are feasible for this application. The rate of evaporation depends on humidity, wind velocity and solar energy absorbed. Dyes may be added to the wastewater to increase the energy absorption, with a consequent increase in the rate of evaporation. Land is a major cost, and problems related to final disposition of the concentrated wastes may arise. Biological and slow air oxidation of the organics may occur. Volatile and odoriferous components must be removed from the wastewater prior to its evaporation.

Disposal and containment. Wastewater can be "controlled" with a minimum of treatment by some disposal or containment options. These options include processed shale wetting as part of the disposal procedure. The water and contaminants are either "cemented" or adsorbed into the processed shale. Provision of an impermeable lining under the shale pile can prevent water from percolating through to the ground if the shale does not cement. Water used for processed shale wetting should not contain any volatiles. Since water used for revegetation and leaching of processed shale piles will contribute to runoff, it may have to be of considerably higher quality than that used for moistening.

Wastewater may be injected underground (deep well injection), as in disposal of some oil well brine wastes (Mercer, Campbell and Wakayima, May 1979). However, costs for underground injection may be significant because deep wells are required to prevent contamination of upper level aquifers. Legal and environmental problems associated with underground injection have not been clarified. Reinjection of mine drainage waters may be a possibility for disposal of this stream when excesses exist. Geologic and hydrologic effects may require evaluation.

Control Technologies Analyzed--

The primary stream which may require control of dissolved organics is:

- Excess Mine Water (stream 75).

Aeration of the excess mine water by bubbling air through it was examined as a dissolved organics control technology. Aeration serves many purposes; for example, it provides oxygen for biological activity in the water, carries out oxidation of chemically oxidizable organics, oxidizes some inorganics and removes odorous compounds. Two examples, reflecting slight changes in the water distribution in the plant, were analyzed to obtain the cost and design information for the treatment, as presented in Table 5.2-21. A cost curve for the aeration pond is shown in Figure 5.2-20.

TABLE 5.2-21. DESIGN AND COST OF AERATION POND

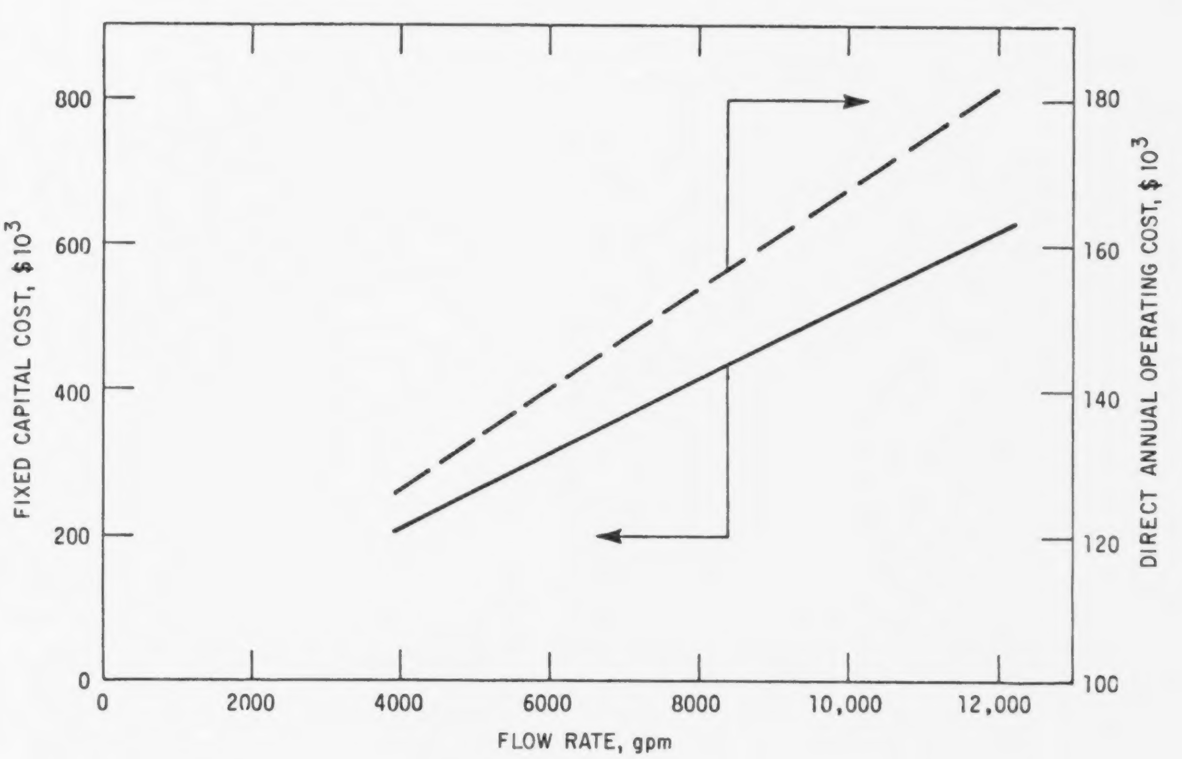
Item	Unit	Example I	Example II[a]
Excess Mine Water Rate	gpm	8,330	8,149
Retention Time	day	1	1
Pond Depth	ft	10	10
Surface Area	10^3 ft^2	160	157
Capacity of Aerator	ft^3/min. of air	7,950	7,340
Fixed Capital Cost	\$$10^3$		
Land preparation		224	220
Aerators		206	190
TOTAL		430	410
Direct Annual Operating Cost	\$$10^3$		
Maintenance @ 4%[b]		14	14
Labor, 10 hr/day @ \$30/hr		99	99
Electricity @ 3¢/kW-hr		40	37
TOTAL		153	150
Total Annual Control Cost[c]	\$$10^3$	262	--

[a] In Example II, more of the mine water is used for processed shale moisturizing; therefore, a lower amount is available for treatment and disposal.

[b] Maintenance is based on the fixed capital cost less contingency.

[c] See Section 6 for details on computation of the total annual control cost. No cost is given for Example II as it is not part of the case study.

Source: WPA estimates.



SOURCE: WPA

FIGURE 5.2-20 COST OF AERATION POND

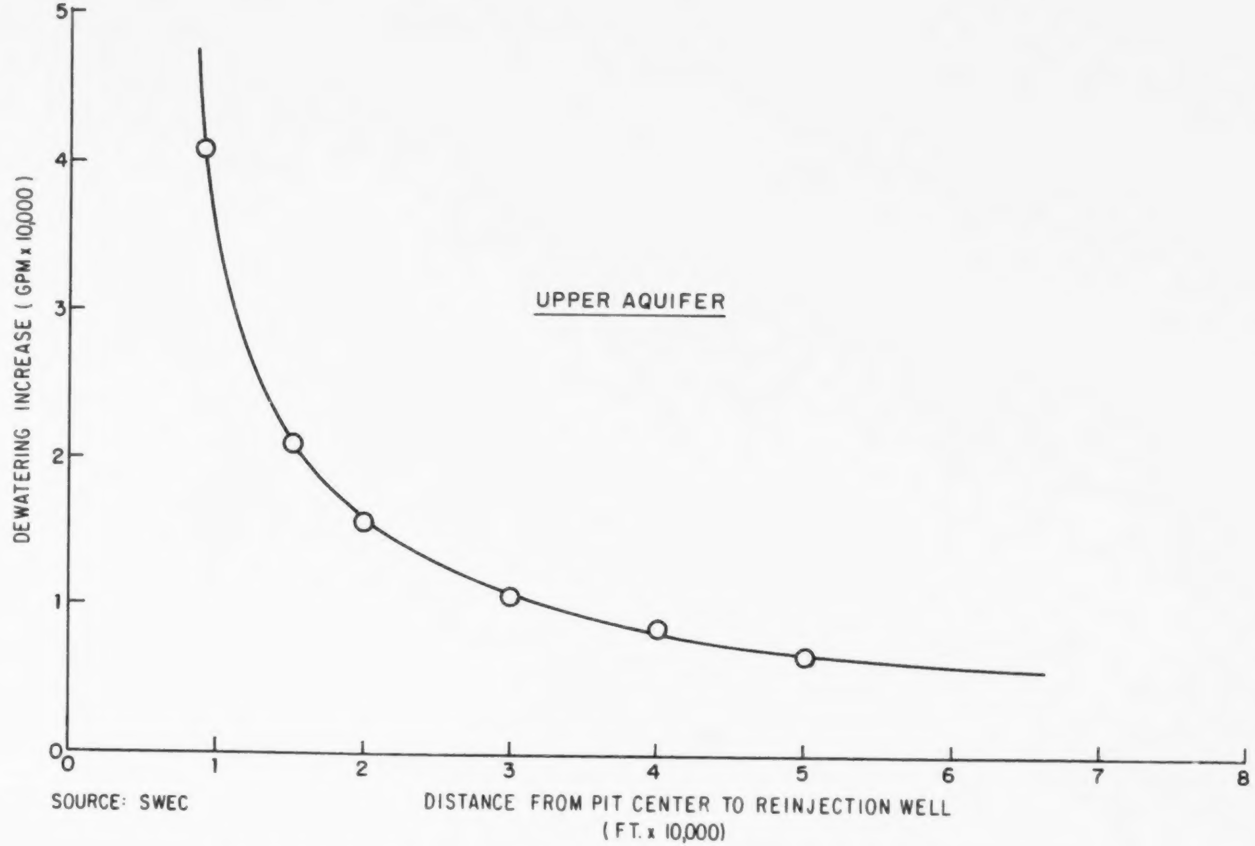
Other Control Technologies Analyzed--

Reinjection of the excess mine water back into the aquifers was analyzed as a viable alternative to surface discharge. This approach has been mentioned for Tract C-a in the event that excess mine water remained after the process needs (Gulf Oil Corp. and Standard Oil Co. [Indiana], March 1976). A combined dewatering rate of 16,500 gpm was calculated from the published data for the two aquifers under the tract, and approximately 8,300 gpm of the mine water were estimated to remain after fulfilling the process requirements. This value was used in determining the essential criteria for reinjection.

The reinjection option has an interesting feature built in: that is, reinjection of the excess mine water back into the aquifers will increase the flow at the dewatering wells. Even more water will now be available for reinjection which, in turn, will again increase the dewatering rate. The extent of the flow increase is dependent upon the reinjection distance from the pit--the farther the reinjection point, the smaller the influence on dewatering. The increases in dewatering rates at equilibrium, as a function of the distance from the pit center, have been determined by an iterative process for the two aquifers and are presented in Figures 5.2-21 and 5.2-22. Figure 5.2-23 represents reinjection pressure as a function of distance. A distance of 50,000 feet from the pit center was finally selected for the reinjection into the upper aquifer after taking into consideration the pressures, flow increases, etc., involved. At equilibrium, approximately 15,000 gpm of the excess mine water will need to be reinjected, causing a flow-back of 7,000 gpm at the dewatering wells, for a total dewatering rate of 23,500 gpm. The design and cost details for the reinjection system are given in Table 5.2-22, and a cost curve is shown in Figure 5.2-24.

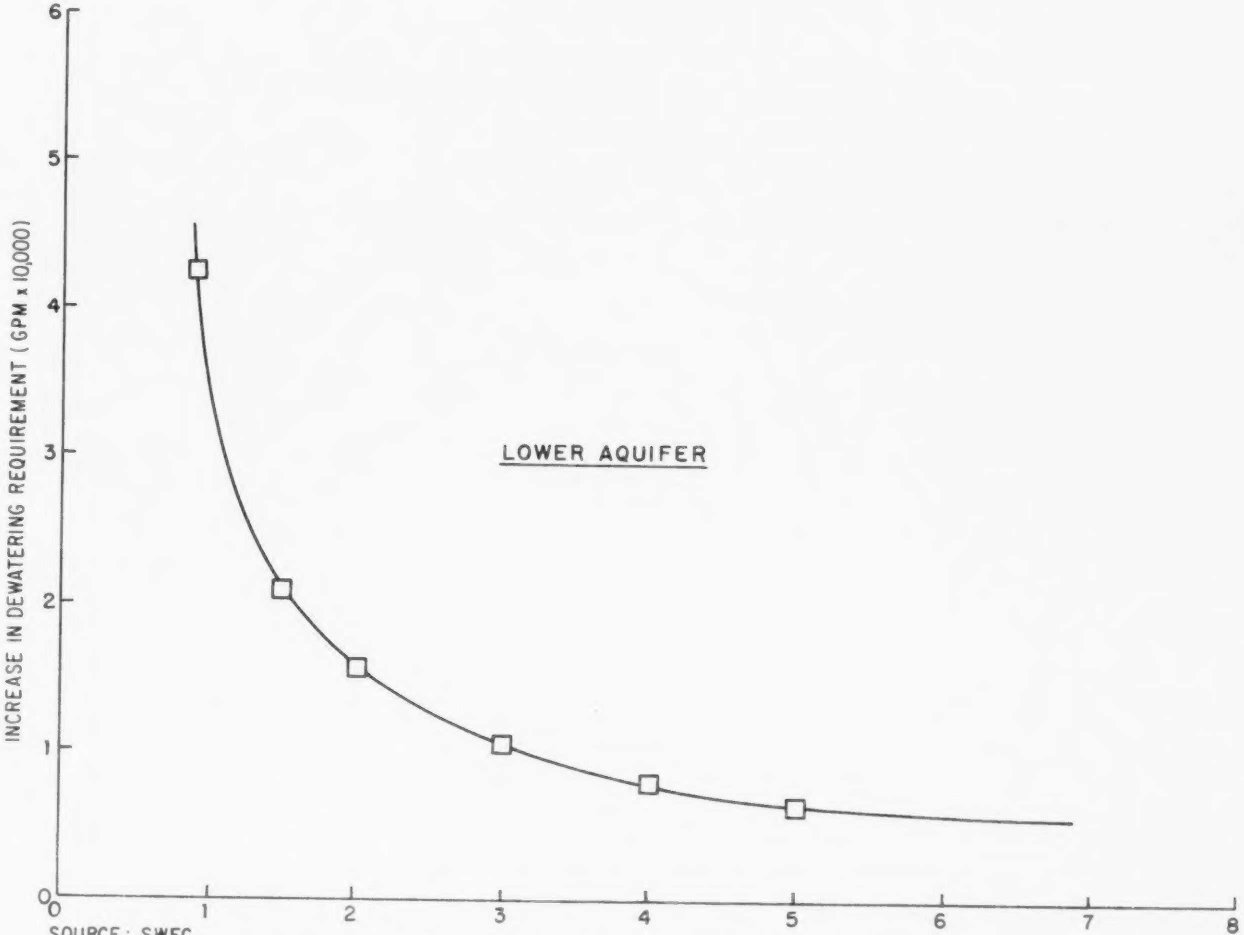
If the use of wastewaters with high organics loading is not acceptable for processed shale moisturizing or reuse in the plant, additional organics removal efficiency can be achieved by several technologies, such as reverse osmosis and carbon adsorption. These technologies have not been proposed for the Lurgi-Open Pit plant, but they have been analyzed based on their potential for application in oil shale wastewater treatment.

Reverse osmosis affords simultaneous removal of the dissolved inorganics and organics. This technology has already been discussed under Dissolved Inorganics control. Under optimum conditions, high removal of dissolved compounds is obtainable with RO, but the permeate from RO may still contain some low molecular weight organic compounds. This stream can be subjected to organics polishing by adsorption on activated carbon. With this technology, the wastewater is allowed to pass through a bed of activated carbon on which the dissolved organics are adsorbed and a cleaner water emerges. The spent carbon is regenerated periodically by steam or hot gas stripping, and the desorbed material is incinerated before it is vented to the atmosphere. If the bulk organics and inorganics have been removed previously (e.g., by RO treatment), the carbon adsorption treated water can be used for high quality water needs (e.g., as a makeup to the cooling tower). Figure 5.2-25 shows the process flow diagram for carbon adsorption (a flow scheme for the technology, when applied to the gas liquor, was already presented in Figure 5.2-17). Table 5.2-23 indicates the composition of the treated water,



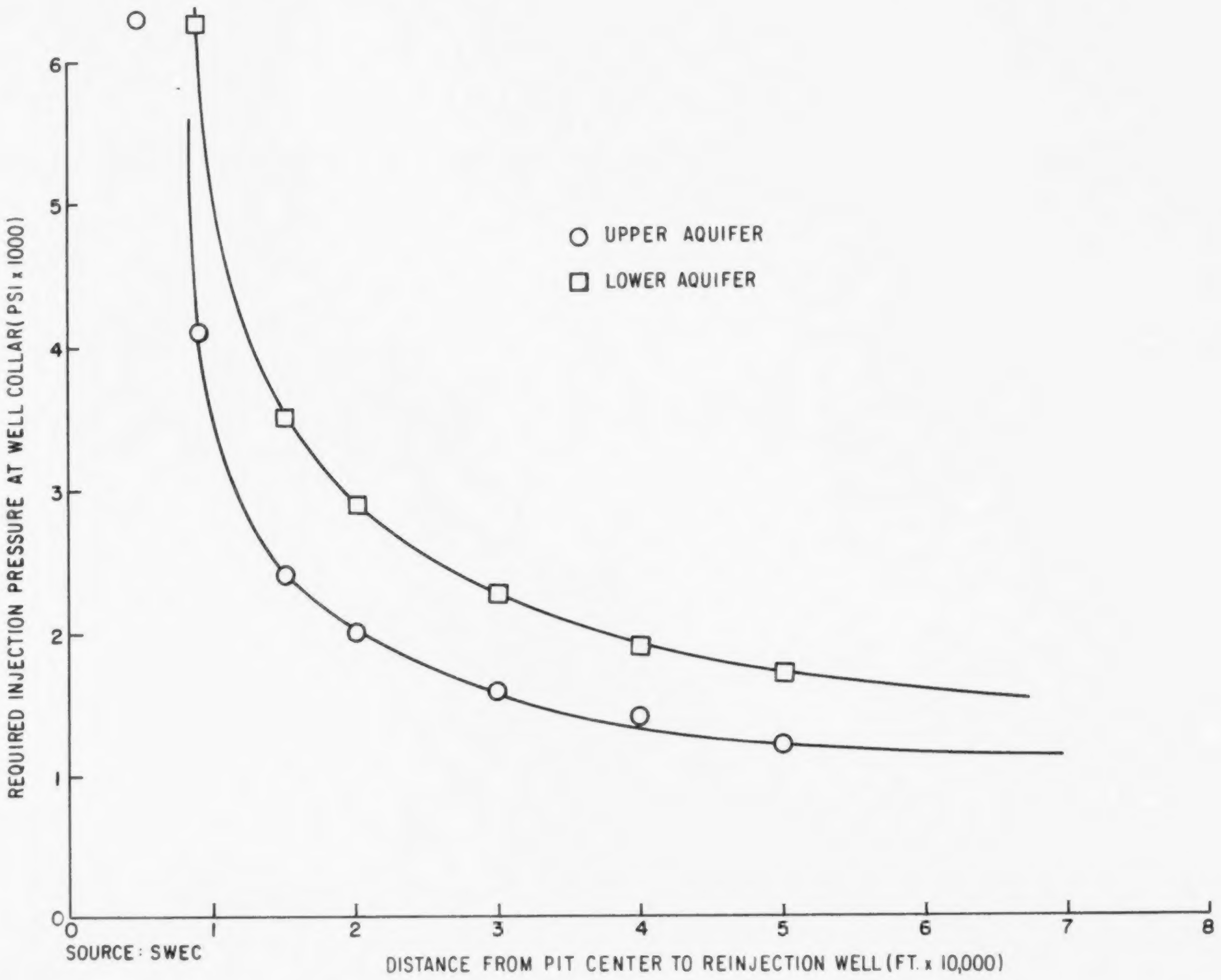
SOURCE: SWEC

FIGURE 5.2-21 UPPER AQUIFER DEWATERING RATE INCREASE WITH REINJECTION DISTANCE



SOURCE: SWEC

FIGURE 5.2-22 LOWER AQUIFER DEWATERING RATE INCREASE WITH REINJECTION DISTANCE



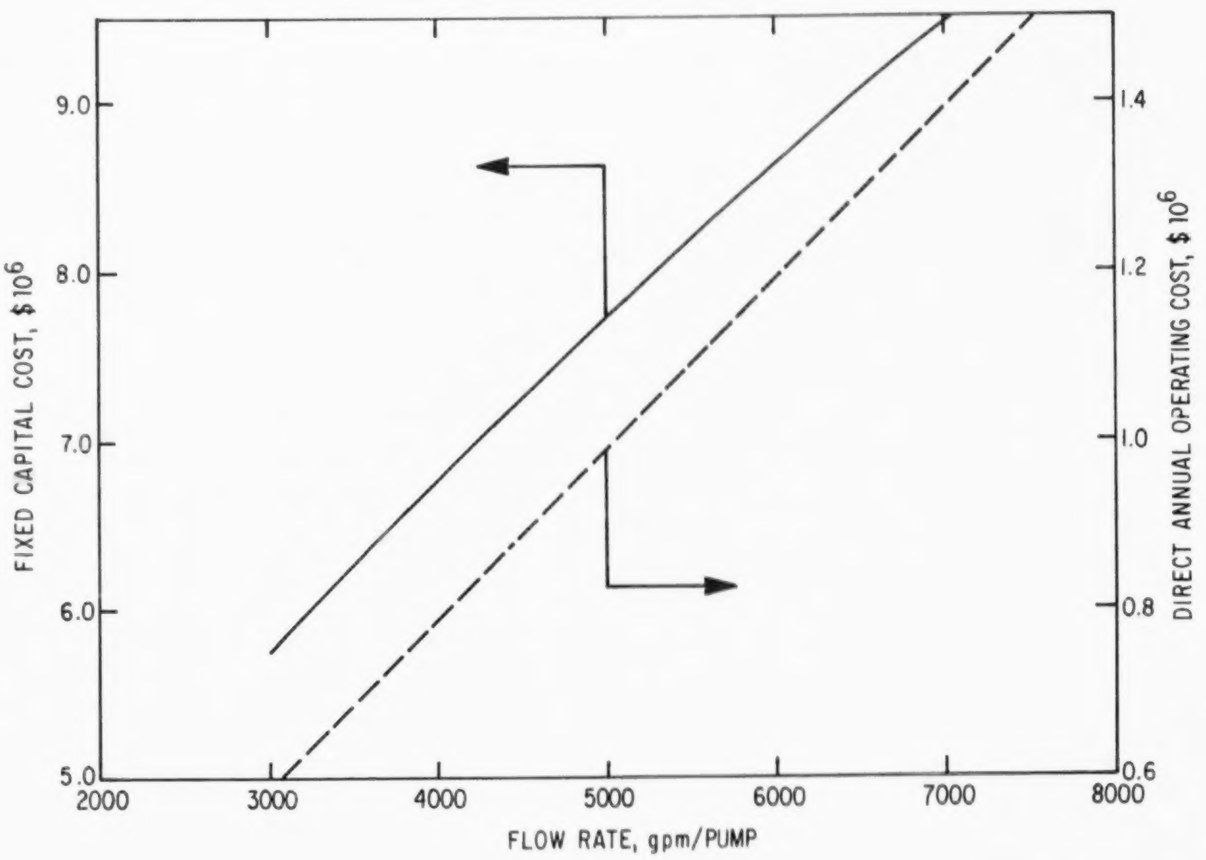
SOURCE: SWEC

FIGURE 5.2-23 REINJECTION PRESSURE AS A FUNCTION OF DISTANCE

TABLE 5.2-22. DESIGN AND COST OF REINJECTION SYSTEM

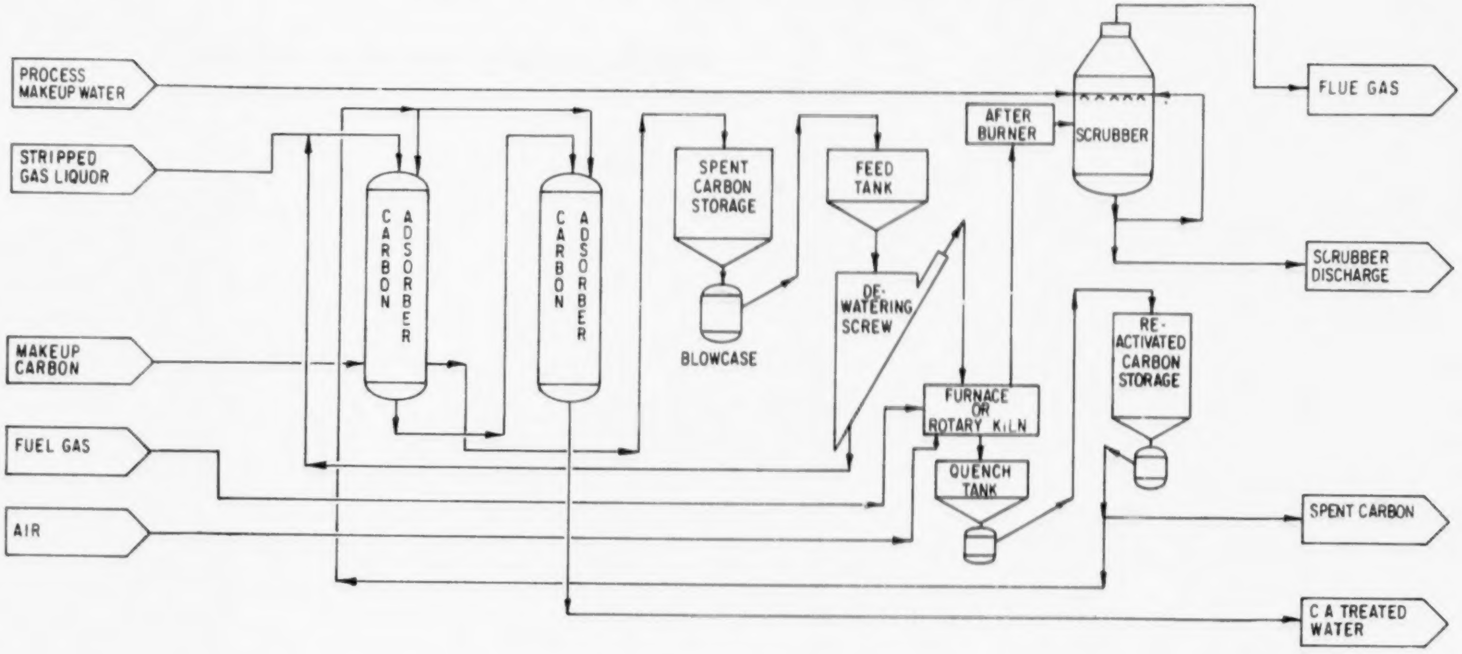
Item	Unit	Quantity
Excess Mine Water Flow Rate	gpm	15,330
Pipeline Pumps	--	3
Flow rate (each)	gpm	5,100
Capacity (each)	gpm	7,500
Discharge pressure	psig	150
Motor (diesel driven)	HP	1,000
Carbon Steel Pipe		
Length	ft	50,000
Diameter	in	30
Design pressure	psig	200
Insulated Carbon Steel Pipe		
Length	ft	5,000
Diameter	in	10
Design pressure	psig	1,500
Reinjection Pumps	--	30
Flow rate (each)	gpm	510
Capacity (each)	gpm	750
Discharge pressure	psig	1,200
Motor (diesel driven)	HP	750
Reinjection Wells	--	10
Carbon steel casing diameter	in	10
Depth	ft	450
Design pressure	psig	1,500
Valves	--	5
Diameter	in	30
Valves	--	60
Diameter	in	10
Diesel Storage Tank		
Capacity	gal	50,000
Fixed Capital Cost	$\$10^3$	
Pipeline pumps		160
Pipe (30")		15,620
Pipe (10")		655
Reinjection pumps		5,853
Reinjection wells		685
Valves (30")		79
Valves (10")		308
Diesel tank		49
TOTAL		23,409
Direct Annual Operating Cost	$\$10^3$	
Maintenance		123
Utilities		2,898
TOTAL		3,021

Source: SWEC estimates.



SOURCE: DRI based on information provided by SWEC

FIGURE 5.2-24 COST OF EXCESS MINE WATER REINJECTION


PROCESS MAKEUP WATER
STRIPPED GAS LIQUOR
MAKEUP CARBON
FUEL GAS
AIR
CARBON ADSORBER
CARBON ADSORBER
SPENT CARBON STORAGE
BLOWCASE
FEED TANK
DE-WATERING SCREW
FURNACE OR ROTARY KILN
QUENCH TANK
AFTER BURNER
SCRUBBER
RE-ACTIVATED CARBON STORAGE
FLUE GAS
SCRUBBER DISCHARGE
SPENT CARBON
C A TREATED WATER

STREAM IDENTITY	FUEL GAS	STRIPPED GAS LIQUOR	MAKEUP CARBON	AIR	SCRUBBER DISCHARGE	SPENT CARBON	C A TREATED WATER	SCRUBBER FLUE GAS	PROCESS MAKEUP WATER
FLOW RATE 10^3 ACFM 10^3 lb/hr gpm	N.D.	281.6 562	0.04	N.D.	N.D.	0.04	281.6 562	N.D.	N.D.
TEMP., °F	AMB	110	AMB	AMB	N.D.	N.D.	AMB	N.D.	AMB
PRESS., psig	N.D.	AMB ←							← AMB

SOURCE: WPA

FIGURE 5.2-25 CARBON ADSORPTION PROCESS FLOW SCHEME

Table 5.2-24 gives the design specifications and cost information for the carbon adsorption technology, and Figure 5.2-26 presents a cost curve for the technology.

TABLE 5.2-23. MATERIAL BALANCE AROUND CARBON ADSORPTION UNIT

Component	Before Treatment		After Treatment	
	Mass %	lb/hr (gpm)	Mass %	lb/hr (gpm)
NH_3	0.0021	6	0.0021	6
$(NH_4)_2SO_3$	0.15	429	0.15	429
Organics (TOC)	0.06	170	0.03	85
H_2O	99.79	281,049 (562)	99.82	281,049 (562)
TOTAL	100.00	281,654	100.00	281,569

Source: WPA estimates.

5.2.5 Water Requirements

Steam Production--

Approximately 1 million lb/hr of 550 psig steam are produced by waste heat recovery in the Lurgi retorting system. The steam is of high quality because only clarified mine water is used. A small portion of the high pressure steam is reduced to 60 psig by driving the retort gas compressor turbines. The low pressure steam thus generated is circulated to various areas of the plant to meet other requirements. This low pressure steam condenses upon use and is returned to the boilers without treatment. Since a large portion of the high pressure steam is not used, it is available for power generation.

Table 5.2-25 presents the steam balance for the plant; as indicated, approximately 866,000 lb/hr, or over 80%, of the total steam is available as a net product. This amount is equivalent to 120 MW of electricity. The power requirement for the lift pipe air compressor is estimated to be about 150 MW; thus, the excess steam can satisfy about 80% of this requirement.

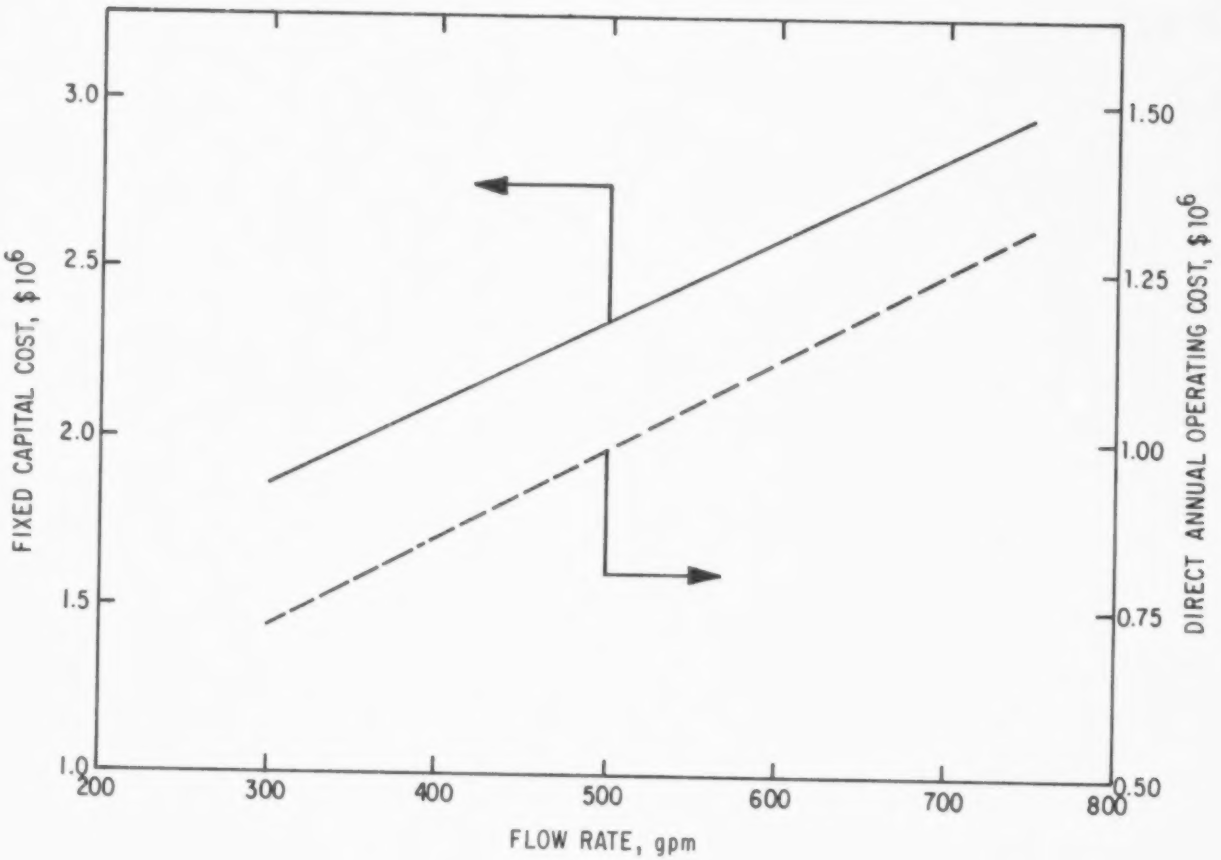
A 0.5% loss factor and 1% blowdown is assumed for the total steam produced. This loss is made up with additional clarified water. Both the feedwater and the makeup water undergo boiler feedwater treatment by zeolite softening and demineralization. Estimated water quality parameters for the boiler feedwater are indicated in Table 5.2-26.

TABLE 5.2-24. DESIGN AND COST OF ACTIVATED CARBON ADSORPTION FOR PROCESS WATERS

Item	Unit	Quantity
Stripped Gas Liquor Flow Rate	gpm	562
Organic Loading	mg COD/l	1,600
Organics Removed	lb COD/hr	800
Carbon Capacity	lb COD/lb C	0.6
No. of Beds (1 standby)	--	2
Bed Diameter	ft	12
Bed Depth	ft	6.5
Carbon Volume/Bed	ft^3	3,350
Carbon Regeneration	lb/day	18,000
Regeneration Period	days	1
Carbon Loss in Regeneration (5%)	lb/day	900
Furnace Area	ft^2	180
Fuel	Btu/lb C	3,000
Steam	Btu/lb C	1,450
Fixed Capital Cost	$\$10^3$	2,500
Direct Annual Operating Cost	$\$10^3$	
Maintenance @ 4%*		81
Labor, 12 hr/day @ $30/hr		118
Regeneration and carbon replacement		882
TOTAL		1,081

* Maintenance is based on the fixed capital cost less contingency.

Source: WPA estimates based on information from Cheremisinoff and Ellerbusch, 1978.



SOURCE: WPA

FIGURE 5.2-26 COST OF ORGANICS CONTROL WITH CARBON ADSORPTION

TABLE 5.2-25. STEAM PRODUCTION, USES AND BOILER FEEDWATER NEEDS

Parameter	Unit	Quantity
Steam Production	10^3 lb/hr	
Waste Heat Boiler		1,060
Steam Uses	10^3 lb/hr	
Ammonia Recovery		53
Stretford Gas Treatment		1
Naphtha Recovery		10
DEA Treatment		130
Net for Power Generation		866
TOTAL		1,060
Net Steam Circulated	gpm	2,120
Feedwater Makeup Requirements	gpm	
Losses (0.5% of circulated)		11
Blowdown		21
Softener Regeneration Waste		11
TOTAL FEEDWATER MAKEUP		43

Source: WPA estimates.

TABLE 5.2-26. WATER QUALITY PARAMETERS FOR BOILER FEEDWATER

Parameter	Low Pressure 0-300 psi	High Pressure 600-750 psi
TDS, mg/l	2,300*	1,300*
Total Alkalinity, mg/l $CaCO_3$	470*	270*
Total Hardness, mg/l $CaCO_3$	0.3	0.2
Iron, mg/l Fe	0.1	0.025
Copper, mg/l Cu	0.05	0.02
Silica, mg/l SiO_2	100*	20*
Specific Conductance, µmhos/cm	4,700*	2,700*

* For a boiler concentration factor of 1.5.

Source: WPA estimates based on data from Krisher, August 28, 1978.

Cooling Water--

Typical cooling water requirements for the Lurgi-Open Pit plant are summarized in Table 5.2-27. Treated mine water could be used as the makeup to the cooling tower. The water quality parameters for the cooling water are indicated in Table 5.2-28. The cycles of concentration are kept low; the relatively large amount of blowdown is used, after equalization with other streams, for processed shale quenching and moistening. Sulfuric acid is added to the makeup water to control carbonate scaling.

TABLE 5.2-27. PLANT COOLING WATER REQUIREMENTS

Water Use	Unit	Quantity
Evaporation	gpm	
Second and Third Condensation Towers		325
Naphtha Recovery		5
Gas Compression		8
Amine Absorber		66
Stretford Gas Treatment		2
Ammonia Recovery		27
Steam Condensing, Plant Drives		450
TOTAL EVAPORATION		883
Cooling Tower Drift (1% of evaporation)	gpm	9
Blowdown	gpm	1,784
TOTAL COOLING TOWER MAKEUP		2,676
Cycles of Concentration	--	1.5

Source: WPA estimates.

Processed Shale Moistening--

The hot processed shale leaving the Lurgi retorting area must be cooled and moistened with water in the processed shale moisturizing mixer before being sent to the disposal area. The hot shale is first quenched, resulting in evaporation of approximately 1,984 gpm of water. The steam generated from the quenching operation is combined with the Lurgi flue gas before entering the electrostatic precipitator. The quenched shale is then moisturized to a final moisture content of approximately 19% to facilitate compaction and stabilization. The optimum moisture content and the extent to which the wastewaters should be treated have not yet been determined. The blowdowns from the cooling tower, boilers, and clarifiers could be used for quenching and moistening. These water streams should not contain volatile material

TABLE 5.2-28. WATER QUALITY PARAMETERS FOR COOLING TOWER RECIRCULATION[a]

Parameter	Limits Minimum	Limits Maximum	Remarks
Langelier Saturation Index[b]	+0.5	+1.5	Nonchromate treatment
Ryznar Stability Index	+6.5	+7.5	Nonchromate treatment
pH	6.0	8.0	
Calcium, mg/l as $CaCO_3$	20-50	300 400	Nonchromate treatment Chromate treatment
Total Iron, mg/l		0.5	
Manganese, mg/l		0.5	
Copper, mg/l		0.08	
Aluminum, mg/l		1	
Sulfide, mg/l		5	
Silica, mg/l		150 100	For $pH < 7.5$ For $pH > 7.5$
$(Ca)\cdot(SO_4)$, product		500,000	Both calcium and sulfate expressed as mg/l $CaCO_3$
TDS, mg/l		2,500	
Conductivity, micromhos/cm^3		4,000	
Suspended Solids, mg/l		100-150	
TOC mg/l		600	
NH_3 mg/l		100	
CN^- mg/l		5	

[a] Concentration in makeup obtained by dividing values above by cycles of concentration.

[b] The limits for the Langelier Saturation Index (an indication of $CaCO_3$ saturation) presume the presence of precipitation inhibitors in nonchromate treatment programs. In the absence of such additives, the limits would be reduced to 0 and 0.5.

Source: WPA estimates based on data from Hart, June 11, 1973.

which would be released upon contact with the hot shale. Table 5.2-29 indicates the water flow rates (gpm) for quenching and moisturizing.

TABLE 5.2-29. WATER REQUIREMENTS FOR PROCESSED SHALE DISPOSAL AND DUST CONTROL

Water Use	Water Required Mass % of Shale	Shale Rate 10^3 lb/hr	Water Rate gpm
Processed Shale Disposal			
Quenching	12.5	7,913*	1,984
Moistening	23.0	7,913	3,640
Processed Shale Dust Control	2.9	7,913	459
Revegetation	4.1	7,913	649
Raw Shale Dust Control			
At Mine	3.2	9,916	634
Crushing	1.4	9,916	285
At Plant	1.0	9,916	190

* Dry processed shale rate.

Source: WPA estimates.

Processed Shale Disposal--

At the disposal area, water is needed for dust suppression and for revegetation. Table 5.2-29 also includes the water requirements for these needs. The water required for dust control is 2.9 mass percent of the dry processed shale rate, and the requirement for revegetation is 4.1 mass percent. Any water used in revegetation at the disposal area should be of a quality acceptable for agricultural use.

Dust Control--

The water requirements for mining, crushing, and fugitive dust control are also summarized in Table 5.2-29. These requirements are given as flow rates (gpm), as well as mass percents of the raw shale rate. The mass percents are 3.2%, 1.4%, and 1.0% for mining, crushing, and fugitive dust control, respectively.

Water used in confined mining operations should be low in volatile or toxic materials because mining personnel will be directly exposed to it. Also, the water should contain low amounts of suspended and dissolved solids to reduce clogging and scaling in spray nozzles. The water used in mining, crushing, and fugitive dust control operations cannot be recovered.

Miscellaneous Requirements--

These include potable and sanitary needs, as well as service and fire water requirements. Table 5.2-30 summarizes these water requirements in terms of makeup, discharge and overall water consumption. Any treatment necessary for these waters is standard practice and not a pollution control activity and, therefore, is not discussed in depth.

TABLE 5.2-30. POTABLE AND SERVICE WATER REQUIREMENTS

Water Use	Usage gal/Man-Shift	Consumption %	Employees No.	Makeup gpm	Discharge gpm
Sanitary/Potable					
At Plant	33	28	950	16	10
At Mine	33	28	580	10	8
Service/Fire Water					
At Plant	66	33	950	29	19
At Mine	50	100	580	14	--

Source: WPA estimates.

5.3 SOLID WASTE MANAGEMENT

The Lurgi-Open Pit processing facility will be a source of large quantities of plant wastes which will require disposal. Table 5.3-1 indicates the makeup of the waste material that will be discarded from the plant over a period of 20 years (project life). Sections 3 and 4 give information about the origin and composition of these streams.

The waste material disposal approach and the practices used in the disposal can have a long-lasting impact on the atmosphere and hydrology of the area as well as on the local aesthetics and habitat. The primary areas of environmental concern in this regard are:

TABLE 5.3-1. MAJOR WASTES PRODUCED OVER A PERIOD OF 20 YEARS

Stream Number	Stream Description	Quantity, 10^6 tons	Material Quantity as a Percent of Total Waste Quantity
2	Subore	78.21	5.62
3	Overburden	408.00	29.32
28	Blowdown from Waste Heat Boiler	0.83	0.06
29	Processed Shale	623.86	44.84
59	Spent Amine	N.D.*	N.D.
70	Stripped Gas Liquor	22.05	1.58
88	Humidified Air Cooler Blowdown	26.06	1.87
90	Water for Dust Palliatives	61.81	4.44
91	Processed Shale Revegetation Water	25.58	1.84
92	Raw Shale Leachate	N.D.	N.D.
93	Storm Runoff	5.91	0.42
95	Service and Fire Water	0.75	0.05
96	Mine Water Clarifier Sludge	6.50	0.47
102	Treated Sanitary Water	0.71	0.05
103	Sanitary Water Treatment Sludge	N.D.	N.D.
104	Boiler Feedwater Treatment Concentrate	0.43	0.03
105	Cooling Tower Blowdown	44.27	3.18
109	Clarified Mine Water to Processed Shale Moistening	86.41	6.21
111	Aerated Pond Sludge	N.D.	N.D.
TOTAL		1,391.38	99.98

* N.D. = Not determined.

Source: DRI estimates based on information from Gulf Oil Corp. and Standard Oil Co. (Indiana), March 1976, and Rio Blanco Oil Shale Co., February 1981.

- Surface Hydrology
- Subsurface Hydrology
- Surface Stabilization
- Hazardous Wastes.

This section briefly describes the disposal approaches that may be applicable to the wastes produced from an aboveground retorting facility (e.g., Lurgi-Open Pit) involving surface mining of the oil shale. In addition, a discussion of control technologies available to mitigate the potential impacts in the areas mentioned above is presented. The applicability of these technologies should be determined on a site-specific, case-by-case basis. Specific information for the facilities involving underground mining and aboveground retorting can be found in the TOSCO II PCTM, while specific information for the combined Modified In Situ-aboveground retorting operations can be found in the MIS-Lurgi PCTM.

5.3.1 Disposal Approaches

The following discussion applies to the basic methods for handling solid wastes produced by the Lurgi-Open Pit processes. Generally, the mining method, geography and hydrology of the area, and the waste characteristics influence the applicability of a disposal approach. The key features of each approach are summarized in Table 5.3-2. A discussion of the control technologies applicable to these disposal alternatives is presented later in this section.

Landfills--

A landfill basically entails placing the waste material as a compacted fill in a suitable location. The wastes from the processing facility are transported to the disposal site by conveyors or trucks and then hauled to the active portion of the landfill. Usually, the solids are laid down in lifts of 9-18 inches and compacted to a suitable in-place density. The compacted fill may be built with a proper slope to a vertical height of 40-50 feet and then flattened, or benched, to provide a passageway for the disposal equipment and to facilitate runoff collection. The overall landfill can be constructed gradually in this fashion, using a multiple-bench arrangement.

Depending upon the geography of the disposal site, the landfill may be built on a level or nearly level surface, in the head of a valley, or across a valley. The applicable control technologies will vary somewhat with site topography but still will be designed to protect the surface and subsurface waters. Applicable control technologies include runon and runoff catchment ponds, embankments and diversion systems, liners and covers, and revegetation. Provision for structural stability of the fill is also a major consideration.

A surface landfill of some type will need to be included in most oil shale developments. This results from the shale undergoing a volume

TABLE 5.3-2. KEY FEATURES OF SOLID WASTE DISPOSAL APPROACHES

Disposal Approach	Principle	Advantages	Disadvantages
Landfills	Place wastes as fill in a convenient surface location and isolate from the surrounding environment.	Relatively simple placement and isolation of wastes. Does not interfere with production.	Dust and erosion control and reclamation/revegetation are relatively labor-intensive operations. Occupies a significant amount of land surface.
Open Pit Backfill	Place wastes as fill in the inactive parts of the pit.	Decreases size of necessary surface landfill. Restores original contours.	Difficult to isolate the wastes from the surrounding environment. Placement is relatively difficult, complex, and interferes with production.
Hazardous Waste Lagoon	Place hazardous wastes in a lined pond and isolate them.	The oil shale developer can maintain absolute control over the waste disposition.	Design, construction, and reclamation may be complex. Requires a relatively level site.

Source: SWEC.

expansion upon mining, crushing, and processing, which precludes all of the shale being returned to the mine.

Open Pit Backfill--

In many respects, the procedures and technologies used in open pit backfilling would be similar to those used in surface landfills. That is, the wastes would be transported to the pit, compacted, and built up to the desired elevation. Stable slopes must be maintained during the simultaneous production and disposal activities and during reclamation, unless the final contour is level with the ground surface.

Runon and runoff collection systems may be necessary to keep the fill and production areas as dry as possible. Permanent groundwater and leachate collection systems may be impractical because the collected water would need to be pumped to the surface and treated for discharge long after the project is shut down. Use of bottom and side liners may be a consideration to reduce the interaction between any leachate produced and groundwater. Placing the wastes in layers to restore the geologic and hydrologic system may also be a consideration.

The pit may be filled below, level with, or above the surrounding ground surface depending upon the quantity of the waste material, site-specific conditions, development plans for the future and permit requirements. A major advantage of backfilling the open pit is that the original contour of the land surface can be more closely restored. Space requirements for the production and disposal activities may be a limiting factor for backfilling small pits.

Hazardous Waste Lagoon--

A hazardous waste lagoon would be a permitted facility either on the project site or off site. It would likely consist of a lined pond designed to be suitable for the containment of hazardous wastes. The major considerations in the design of such a pond would include a runon diversion system, an embankment, one or two impervious bottom liners with a drained sand layer below or between them, a slurry wall beneath the embankment, a surface seal layer, and provisions for reclamation and revegetation (U.S. EPA, September 1980).

Once the lagoon is filled to its capacity, wick drains could be installed to facilitate evaporation, allowing quicker consolidation of the sludge. Gravel could also be added to aid consolidation. An impermeable surface seal may then be added on top and joined with the bottom liner to isolate the wastes from the surrounding environment. The final aspects would include placing subsoil and topsoil over the seal, followed by revegetation of the surface.

5.3.2 Surface Hydrology Control Technologies

Solid waste management practices in the area of surface hydrology entail the handling of surface waters on and around the disposal facility.

Specifically, surface streams and precipitation are prevented from running onto the waste pile and contaminated waters (runoff, leachate) are kept from mixing with the natural waters.

The technologies discussed below are those that are applicable to a surface landfill, and they are summarized in Figure 5.3-1. The key features of the technologies are highlighted in Table 5.3-3 and a more detailed description with cost data is presented in the text.

Runon Diversion System--

A runon diversion system will generally be needed with any surface landfill to prevent surface water from flowing onto the waste material and becoming contaminated or causing erosion. The system may include ditches, lined channels, conduits, and embankments arranged to direct the flow of surface water around or away from the waste material, and energy dissipators to moderate the impact of the flow.

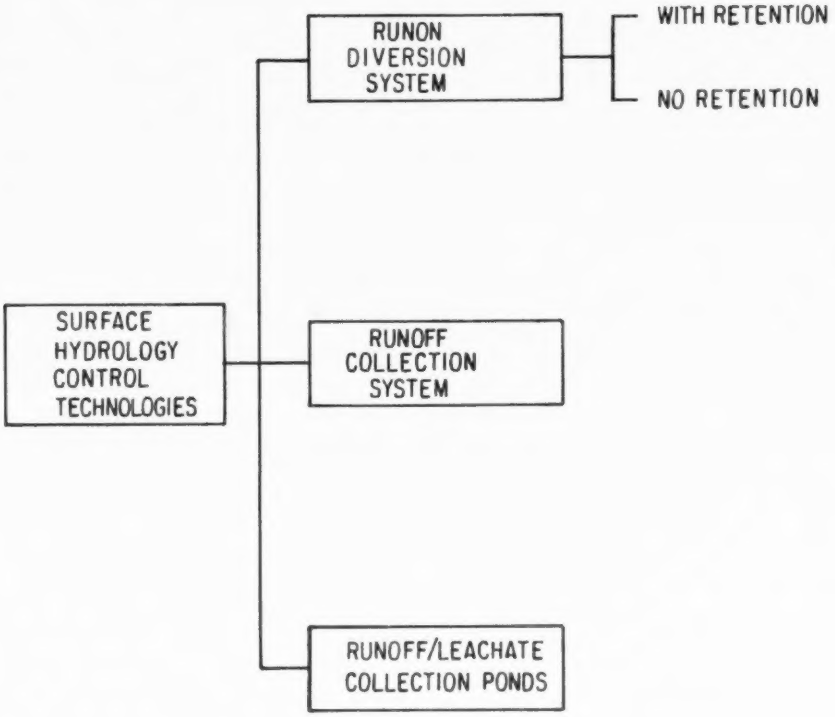
The complexity and extent of the system will vary widely based on the amount of water to be diverted and the arrangement of the site. For a fill on a relatively level site, runon diversion may require only a system of channels and small embankments to deflect surface flow away from the landfill. In the case of a head-of-valley fill or a cross-valley fill, runon diversion might include an embankment dam to retain peak flows from the design storm until they can be passed through a conduit beneath or around the fill. Alternatively, the system may consist of a conduit or channel large enough to pass the design flow without an embankment (without retention).

The design of a runon diversion system will be influenced by: the size of the drainage area and topography which affect the runon rates, retentions, and embankment material quantities; the size, length, and complexity of controlled release structures and channeling systems; and the need for and extent of energy dissipators and/or drop structures. For example, the runon from a site with a large drainage area in a gently sloping topography could be diverted quite efficiently by an unlined canal or channel; another site with small runoff rates, but highly erodible steep topography, may necessitate cost-intensive lined channels, flumes or conduits, as well as drop structures or energy dissipators.

Runoff Collection System--

A runoff collection system usually consists of a system of channels, ditches, and conduits arranged to prevent the surface water that has contacted the waste material from leaving the site. Another purpose of this system is to drain the surface water from the wastes to limit the erosion and infiltration potential. Collected water may also be used to meet process needs.

The basic elements of this system are backsloped benches on the face of the landfill and a means of collecting the water from the fill surface. Generally, half-round pipes, impervious membranes, or highly compacted soil


SURFACE HYDROLOGY CONTROL TECHNOLOGIES
RUNON DIVERSION SYSTEM
WITH RETENTION
NO RETENTION
RUNOFF COLLECTION SYSTEM
RUNOFF/LEACHATE COLLECTION PONDS

SOURCE: SWEC

FIGURE 5.3-1 SURFACE HYDROLOGY CONTROL TECHNOLOGIES

TABLE 5.3-3. KEY FEATURES OF SURFACE HYDROLOGY CONTROL TECHNOLOGIES

Control Technology	Principle	Purpose	Comments
Runon Diversion System	Uses channels and embankments to prevent surface water from contacting the waste material.	Reduces erosion and increases site stability. Reduces the amount of water contacting the waste material, thereby reducing the potential for surface water pollution.	Reduces the amount of water contaminated, thereby reducing treatment costs.
With Retention	Embankment dam holds peak flows for controlled release, evaporation, or percolation into the ground.		Requirements for the channel or conduit are greatly reduced. Provides flexibility in the use of the collected water.
No Retention	Channel or conduit is sized to convey peak flow with no retention of water.		Eliminates the need for runon retention structures and associated maintenance. More expensive than using an embankment for retention and controlled discharge of the peak flow.
Runoff Collection System	Drained benches collect and remove precipitation falling on the disposal site.	Reduces erosion of fill and infiltration into fill. Collects water for reuse or discharge.	Decreases erosion and infiltration. Requires maintenance.
Runoff/Leachate Collection Ponds	Lined ponds are used to retain leachate and runoff.	Prevents release of contaminated waters.	Collects water for reuse, treatment and discharge.

Source: SWEC.

or wastes are used to line ditches which collect the runoff from the bench and the segment of the landfill slope above it, as shown in Figures 5.3-2 and 5.3-3. The ditches empty into central conduits leading to a containment/evaporation pond at the toe of the landfill. On larger piles or in areas with extensive rainfalls, small embankments on the crest of the landfill or on the benches might be used to retain the runoff and thus limit the peak flows into the rest of the drainage system.

A problem with limiting the peak flows using embankments on the waste pile is that the water ponded on the landfill will have a greater tendency to infiltrate the waste material. This increased infiltration could have a detrimental effect on the stability of the slope and will somewhat increase the amount of water which must be handled by the leachate collection system (discussed under subsurface hydrology).

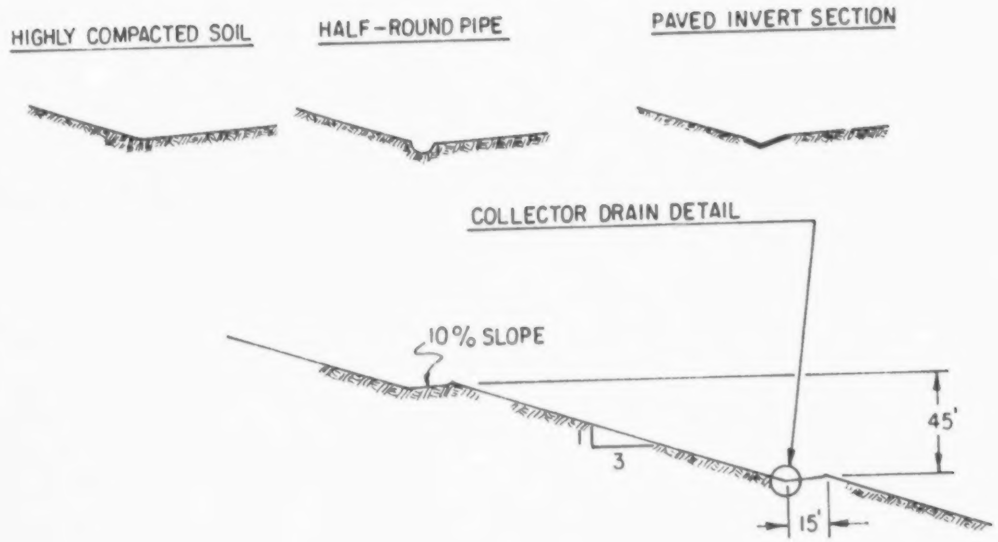
The costs for a variety of runoff collection system designs for surface landfills were estimated and these are plotted in Figure 5.3-4. Example 1 used shaped benches with unlined ditches for lateral conveyance and concrete weir collectors and corrugated metal pipe with energy dissipators for vertical conveyance. It also incorporated some temporary retention of runoff on the waste pile surface, which reduced the necessary capacity and cost of the vertical conveyance portion of the system. Example 2 used split corrugated metal pipe to line the collection ditches to facilitate lateral conveyance, and concrete weir collectors and corrugated metal pipe with energy dissipators for vertical conveyance. Example 3 used the lined ditches for lateral conveyance, with a concrete flume and a stilling basin for vertical conveyance.

The cost data, as can be seen in the plot, are highly dependent on the particular design, and no single cost curve relationship can be drawn through the data points. Example 1, which assumes a more modest design, defines the lower boundary of the cost envelope, and Example 3 defines the high end of the cost envelope.

The design of the runoff collection system for open pit backfills would differ from that for surface landfills because the runoff has to be pumped to the surface for its disposition. Hence, a system for an open pit project might consist of a series of collection sumps located at the junction of the pit wall and landfill, from which the collected water is pumped to the surface and probably used for processed shale moistening. Both the sumps and pumps require only operating expenditures, as any associated capital expenditure is considered to be a part of the mining plan. The total annual operating costs for the sumps and pumps for an open pit mine, as described in Sections 2, 3 and 4, were estimated to be $63,000 and $16,000, respectively, while the total annual control costs were estimated to be $64,000 (0.3 cents/bbl of oil) and $16,000 (0.1 cents/bbl of oil). The details of cost computation are presented in Section 6.

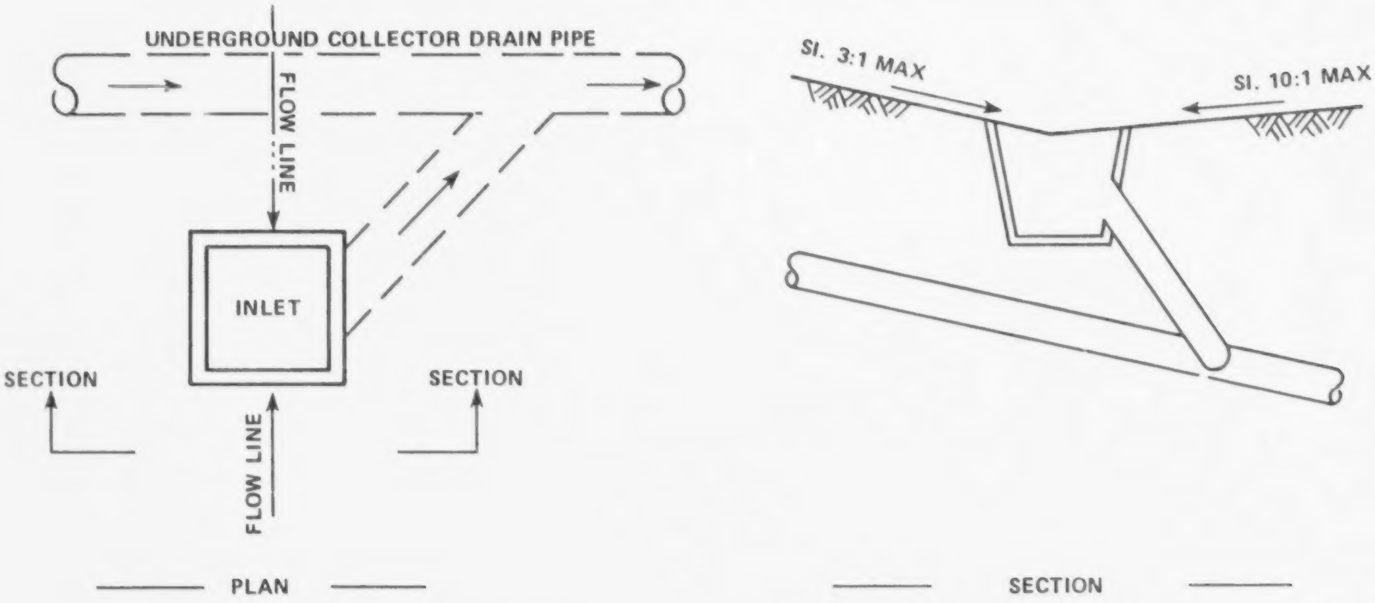
Runoff/Leachate Collection Ponds--

At the outlet of the collection system for surface runoff, a structure is needed to contain the collected water for reuse, treatment and discharge,


HIGHLY COMPACTED SOIL
HALF-ROUND PIPE
PAVED INVERT SECTION
COLLECTOR DRAIN DETAIL
10% SLOPE
1
3
45'
15'

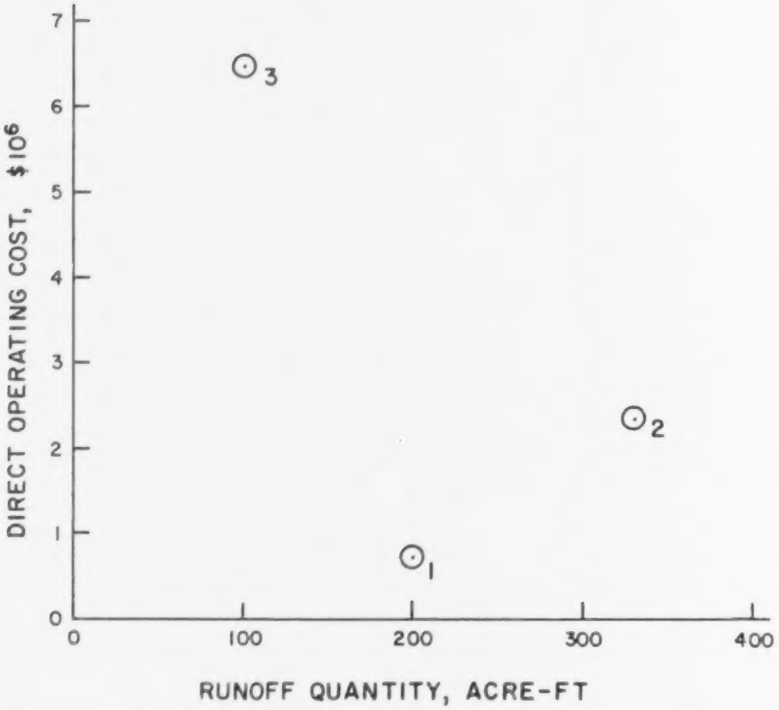
SOURCE: SWEC based on Colony Development Operation, March 1980

FIGURE 5.3-2 TYPICAL RUNOFF COLLECTION SYSTEMS


UNDERGROUND COLLECTOR DRAIN PIPE
FLOW LINE
INLET
SECTION
SECTION
FLOW LINE
PLAN
Sl. 3:1 MAX
Sl. 10:1 MAX
SECTION

SOURCE: SWEC based on Colony Development
Operation, March 1980

FIGURE 5.3-3 RUNOFF COLLECTION AND CHANNELING



NOTES:

Example 1 utilizes shaped benches as unlined ditches for lateral runoff conveyance.

Examples 2 & 3 utilize split corrugated metal pipe to line ditches for lateral runoff conveyance.

Examples 1 & 2 utilize buried corrugated metal pipe and energy dissipators for vertical runoff conveyance.

Example 3 utilizes a concrete flume with stilling basin for vertical runoff conveyance.

The costs indicated are cumulative for the project life.

See Section 6.2.3 for details on the solid waste management cost methodology.

SOURCE: SWEC

FIGURE 5.3-4 RUNOFF COLLECTION COSTS

or for evaporation. The structure would consist of an embankment across a former stream channel to form a pond, and the pond may be lined or unlined depending upon the nature of the impounded material. If a liner is needed, it would be protected from wave action, as necessary, using rip-rap, a sand layer, soil cement or similar materials. Since the pond would be located at the base of the landfill, it might also be used to collect the leachate from the fill.

Cost data for four examples of runoff/leachate collection ponds for surface landfills are presented in Figures 5.3-5 and 5.3-6. Figure 5.3-5 presents the total cost of the embankment and liner as a function of the construction material quantities used in each case, while Figure 5.3-6 isolates the cost of the liner as a function of the liner material quantity only. Examples 1, 2 and 3 utilized compacted processed shale as the liner, while Example 4 used Mancos Shale as the liner. The relatively high cost of using an off-tract material (Example 4) is evident in the figures. The cost increase is incurred due to the source development, processing and hauling of Mancos Shale. Slight cost differences may be observed between similar systems, and these can be attributed to site-specific features, such as the arrangement and configuration of the embankments and ponds.

A runoff collection and containment system for a pit backfilling approach differs from that for the surface landfills. Instead of an embankment and pond downgradient from the landfill, a series of collection sumps and pumps would be used, as discussed under Runoff Collection System.

5.3.3 Subsurface Hydrology Control Technologies

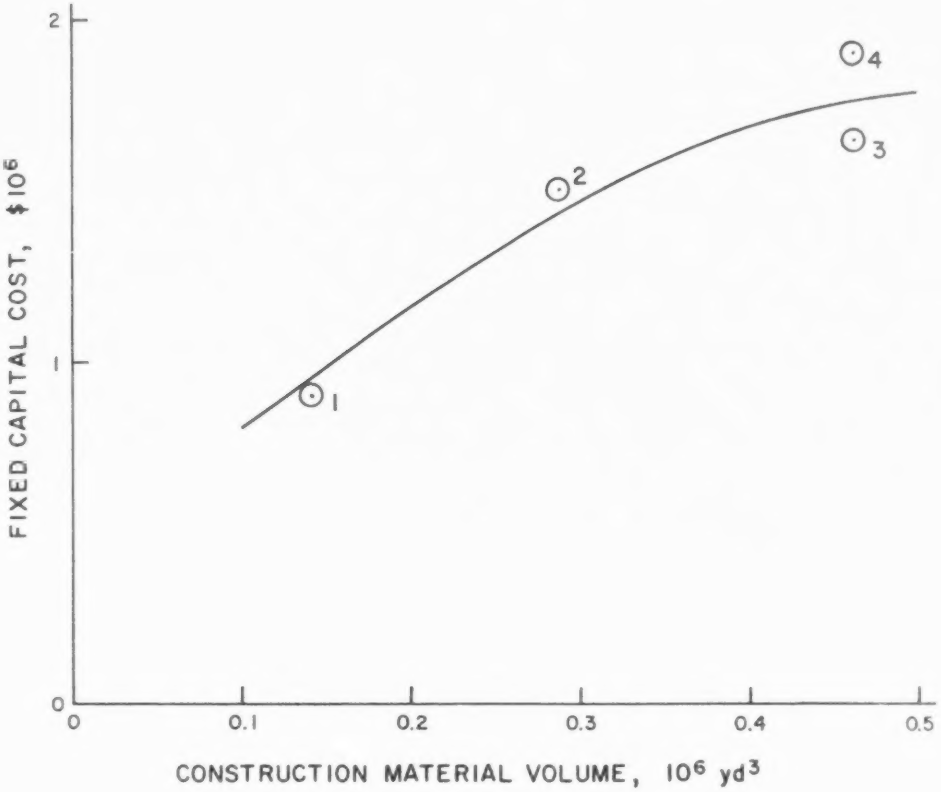
The technologies and practices in the area of subsurface hydrology involve the handling of groundwater seepage under a landfill to prevent infiltration of the pile and the control of water from the pile to prevent contamination of the groundwater. The technologies, as summarized in Figure 5.3-7, are applicable to a surface landfill, and their key features are presented in Table 5.3-4. Detailed descriptions of the technologies, along with cost information, are presented below.

For open pit backfilling, subsurface hydrology control may consist of aquifer dewatering. Since this operation would be an integral part of the mining plan, additional costs for backfilling would not be incurred.

Liners and Covers--

A liner is essentially a material with low water permeability that is installed at the bottom of a landfill or pond. Its purpose is to prevent the contaminated waters from the wastes from mixing with the groundwater. It also prevents groundwater from infiltrating the bottom of the landfill.

A cover is also made up of a low-permeability material and it is used as a surface sealer for the landfill. It prevents the runoff from infiltrating the pile, thereby reducing the quantity of the leachate and minimizing stability problems.



NOTES:

All Examples include cost of embankments and pond liners.

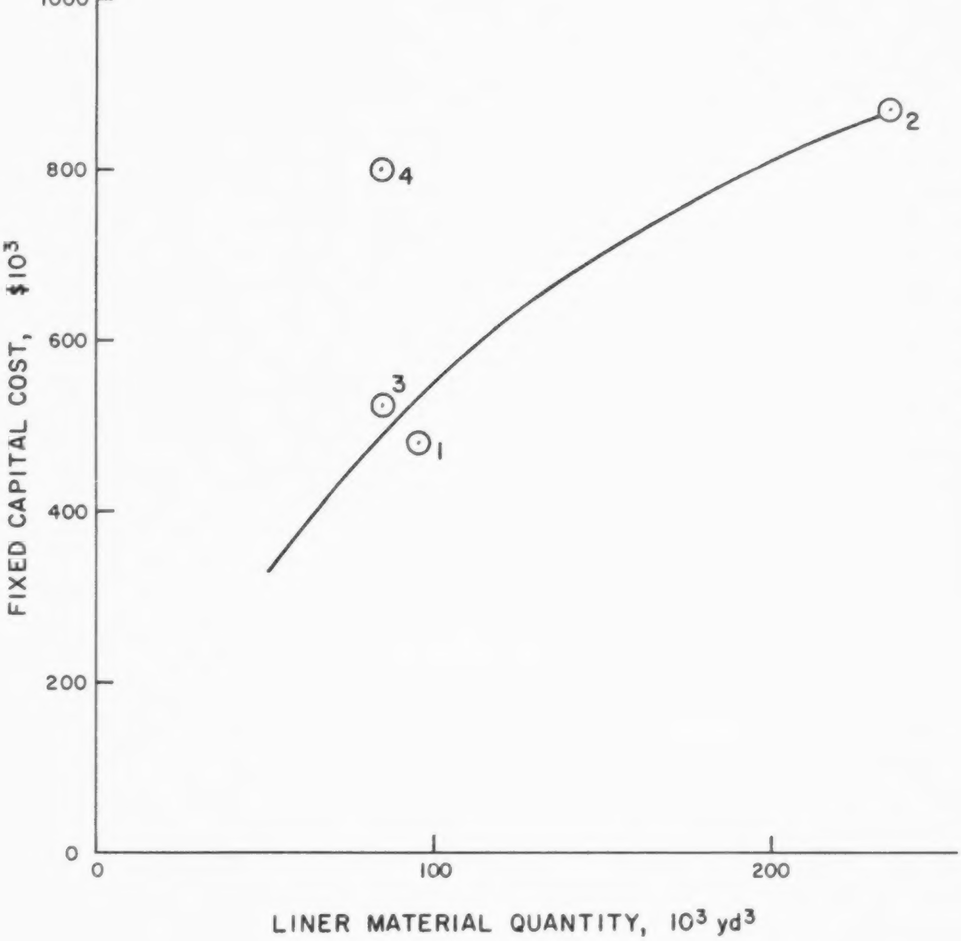
Examples 1, 2 & 3 include pond liners constructed of processed shale.

Example 4 includes a liner constructed of Mancos Shale (off-tract material); cost is increased due to processing and transport.

See Section 6.2.3 for details on the solid waste management cost methodology.

SOURCE: SWEC

FIGURE 5.3-5 RUNOFF/LEACHATE POND COSTS



NOTES:

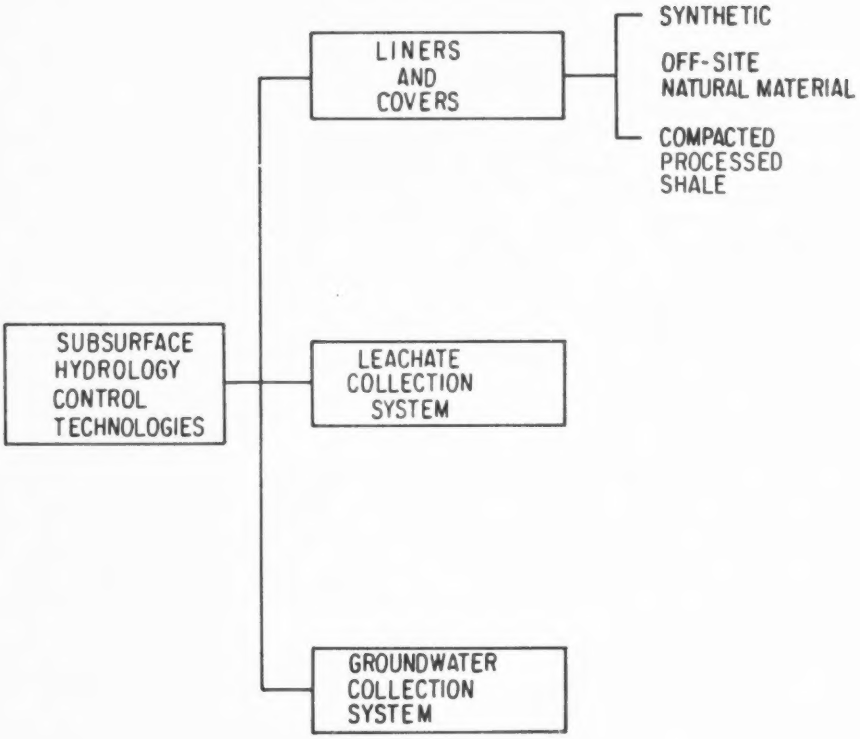
Examples 1, 2 & 3 include liners constructed of processed shale.

Example 4 includes a liner constructed of Mancos Shale (off-tract material); cost is increased due to processing and transport.

See Section 6.2.3 for details on the solid waste management cost methodology.

SOURCE: SWEC

FIGURE 5.3-6 RUNOFF/LEACHATE POND LINER COSTS


SUBSURFACE HYDROLOGY CONTROL TECHNOLOGIES
LINERS AND COVERS
SYNTHETIC
OFF-SITE NATURAL MATERIAL
COMPACTED PROCESSED SHALE
LEACHATE COLLECTION SYSTEM
GROUNDWATER COLLECTION SYSTEM

SOURCE: SWEC

FIGURE 5.3-7 SUBSURFACE HYDROLOGY CONTROL TECHNOLOGIES

TABLE 5.3-4. KEY FEATURES OF SUBSURFACE HYDROLOGY CONTROL TECHNOLOGIES

Control Technology	Principle	Purpose	Comments
Liners and Covers	Low permeability layer severely restricts seepage.	Reduce formation of leachate. Prevent contamination of the groundwater by leachate from the fill. Prevent groundwater invasion of the fill, which might produce instability and additional leachate.	
Synthetic			Provide the lowest permeability but have the highest cost. Long-term durability is questionable.
Off-site Natural Material			High cost. Advantage is long-term durability.
Compacted Processed Shale			Lowest cost. Small particles may infiltrate adjacent drains. Advantage is long-term durability.
Leachate Collection System	Collects leachate at the base of the landfill and drains into the pond.	Reduces groundwater contamination by effectively removing the leachate. Prevents loss of fill stability due to saturation.	Collected water may be used for process needs.
Groundwater Collection System	Collects groundwater seepage beneath the landfill and drain.	Prevents loss of fill stability due to buildup of groundwater pressure beneath the liner.	Collected water may be used for process needs.

Source: SWEC.

There are several materials which can be considered for the liners and covers. Probably the least expensive material would be compacted processed shale. It has the advantage of being readily available at the site. A similar lining could be made of processed shale or clay from off site if the quality of the processed shale from the site is unsuitable; however, these options would be relatively expensive due to the extra handling and hauling costs. There is also a variety of synthetic liners which could be considered. High-density polyethylene, for example, would range upward from a price similar to that for the off-site materials, depending upon the thickness used. This would make it very expensive for use in a processed shale landfill and it may have questionable long-term durability. Another option that could be considered, particularly for a hazardous waste lagoon, is simply a combination of a synthetic liner with one of the other liners mentioned above.

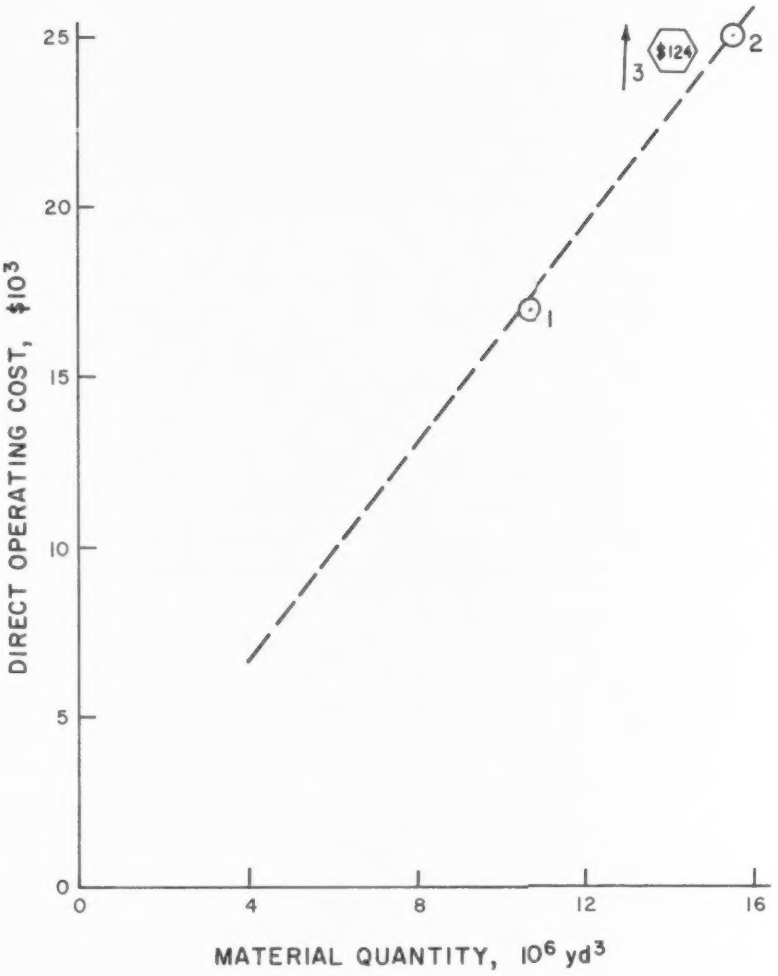
Linings made of natural materials will dry and crack if they are left exposed to the weathering elements for long periods. Therefore, if a pond is not expected to remain at a relatively consistent level, a synthetic liner might be considered. Hazardous waste lagoons sometimes have double liners; however, the catchment and evaporation ponds presumably will need only one liner or no liner since they will not contain hazardous materials. If a combination of two liners is used, the synthetic liner may be placed above the natural material liner to prevent its drying and cracking. In cases where a synthetic liner is used, it should be covered by a layer of sand or gravel to protect it from traffic and wave action. Also, because of the weight of the fill and because the fill may be placed above an underground mine, the liner must accommodate a certain amount of subsidence and stretching and still function properly.

The cost of liners and covers depends on the quantity and type of material used. Figure 5.3-8 presents the costs for three separate liner and cover systems for surface landfills. Examples 1 and 2 assumed the use of highly compacted processed shale for construction of the liners, while Example 3 assumed the use of Mancos Shale. The compacted processed shale represents the lowest material cost option, while Mancos Shale is a more expensive natural material since it has associated source development, processing and hauling costs. The cost curve in the figure may be used to obtain an "order-of-magnitude" estimate of liner cost utilizing highly compacted processed shale as the construction material. The estimated cost for other liner materials would fall above this curve to a degree which is dependent on the source development, processing, and hauling costs associated with delivering these materials to the disposal site.

Leachate Collection System--

The purpose of a leachate collection system is to collect water which infiltrates a landfill and drain it efficiently in order to prevent the saturation of the landfill and contamination of groundwater beneath the waste pile, as well as to facilitate handling of the leachate.

Leachate collection systems typically consist of blankets, or zones, of highly pervious sand and gravel. In some cases this is augmented with



NOTES:

Examples 1 & 2 utilize 3 feet of highly compacted processed shale for liner material.

Example 3 utilizes 3 feet of compacted Mancos Shale (off-tract material) for liner material; cost of processing and hauling this material makes this option more expensive than the others.

The costs indicated are cumulative for the project life.

See Section 6.2.3 for details on the solid waste management cost methodology.

SOURCE: SWEC

FIGURE 5.3-8 LINER COSTS

embedded perforated pipe to increase the capacity, and it may also include collector ditches where the system emerges onto a broad level area. The sand or gravel layer would be located just above the bottom liner and it may be wrapped in filter fabric or surrounded by carefully graded sand filters to prevent infiltration by the processed shale particles. In either case, the collection system should be designed so that movement and settlement do not result in discontinuity of the gravel layer or impede drainage to the collection or evaporation ponds.

The costs for four distinct leachate collection systems for surface landfills were estimated and these are presented in Figure 5.3-9. In Examples 1 and 2, due to the valley shape of the disposal site, only the drain material was necessary for the collection system. The leachate in these two cases was drained in the runoff/leachate collection pond located downstream from the landfill. In Example 3, a toe ditch was necessary to collect the leachate due to the presence of the broad valley area at the toe of the landfill. The ditch was then drained into the common runoff/leachate collection pond. Example 4 also required a toe ditch which was drained into a leachate collection pond, while the runoff was impounded separately in evaporation ponds on the waste pile surface. Examples 3 and 4 required the same drainage material quantity. The cost difference between the two examples is due to the inclusion of a separate collection pond in Example 4. Data point 5 on the figure represents the cost of drainage material only for Examples 3 and 4. The cost of the toe ditch may be obtained by subtracting data point 5 from 4.

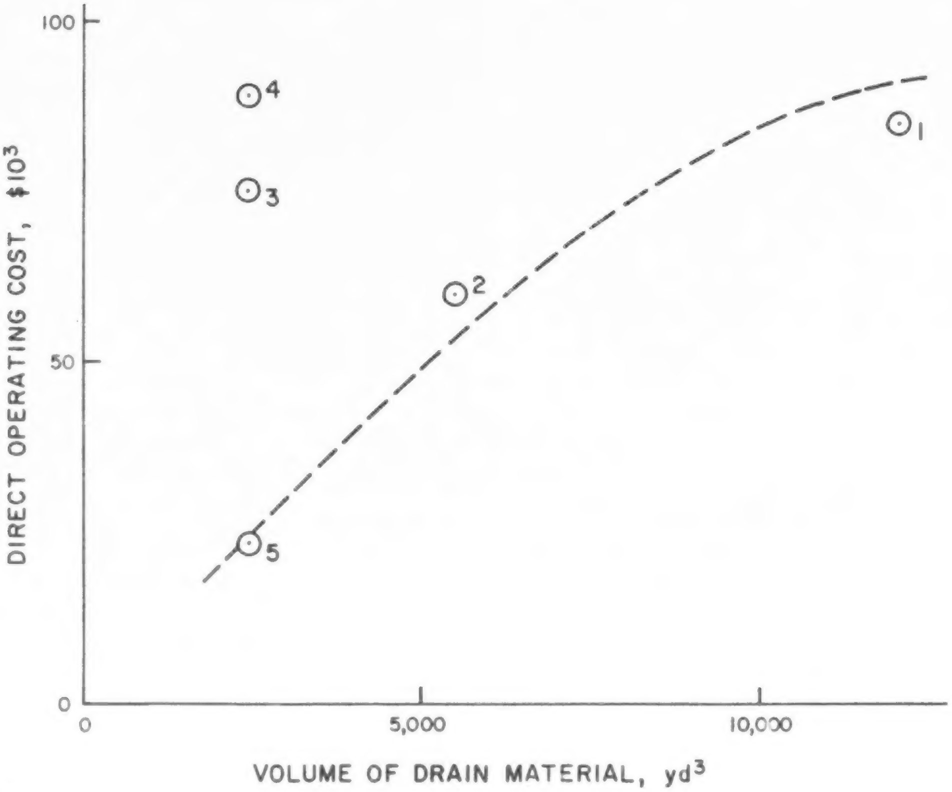
The costs for similar systems should be proportional to the volume of drainage material used, but slight deviations may be encountered due to the site-specific conditions.

For open pit mining and backfilling operations, some leachate is likely to be collected in the pit along with the runoff and it may be used for processed shale moisturizing. Controlling the leachate after the backfilling operations have been completed would not be practical. Therefore, emphasis should be placed on minimizing the production of leachate. Some considerations in this regard would be to reduce the overall permeability of the backfilled mass and to minimize penetration of surface water by utilizing a cover.

Groundwater Collection System--

The purpose of a groundwater collection system is to relieve pressure from the seeps and springs beneath a landfill. This situation is most likely in the cases of cross-valley or head-of-valley landfills. The system will be essentially identical to the leachate collection system except it would be below the bottom liner rather than above it.

Groundwater collection systems typically consist of blankets or zones of pervious sand and gravel drained beyond the perimeter of the landfill. This may be augmented with embedded perforated pipe to increase capacity and with collector ditches. The sand or gravel layer would be lined as necessary with filter fabric or surrounded by properly graded sand filters to prevent



NOTES:

Examples 1 & 2 require only drain material due to the valley shape; leachate containment is performed by the contaminated runoff catchment pond of which the leachate is a negligible component.

Example 3 includes cost of toe ditch for collection due to broad valley at waste pile toe; containment is also by the contaminated runoff catchment pond.

Example 4 includes toe ditch collection and separate containment pond because, in this case, contaminated runoff is contained in evaporation ponds on the waste pile surface.

Example 5 includes only the drain material cost of Examples 3 & 4.

The costs indicated are cumulative for the project life.

See Section 6.2.3 for details on the solid waste management cost methodology.

SOURCE: SWEC

FIGURE 5.3-9 LEACHATE COLLECTION COSTS

infiltration of smaller particles from adjacent materials. The system must also be designed to maintain its continuity despite possible subsidence or settlement of the landfill.

The costs of two groundwater collection systems for surface landfills were estimated and these are plotted in Figure 5.3-10. Both systems used gravel blankets under the pile to collect the groundwater seepage. In Example 2 the gravel blankets were used only above the seeps and springs, while in Example 1 an extensive network of the blankets was considered, resulting in a higher cost. The cost of the collection system should be proportional to the quantity of the drainage material used.

The use of a groundwater collection system under an open pit backfill does not appear practical, especially in areas like Tract C-a where a large amount of groundwater exists. A control over the groundwater flow in the pit during active operations is achieved by dewatering the aquifers, which is performed to keep the pit as dry as possible to facilitate mining; hence, it is not considered a solid waste management technology. At the completion of the project, the dewatering wells are shut down and original groundwater levels are reestablished.

Some conceptual controls, such as hydrologic barriers and bypass, may be applied to reduce the groundwater interaction with the backfilled material. These are discussed in the MIS-Lurgi PCTM.

5.3.4 Surface Stabilization Technologies

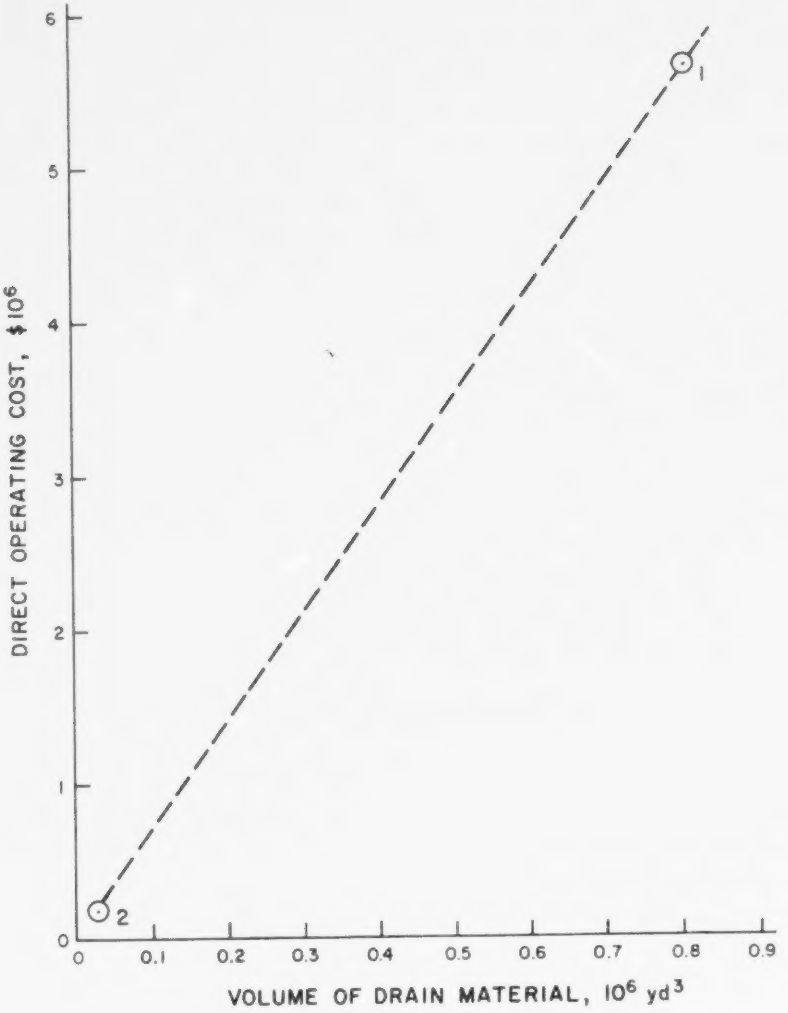
The activities and technologies in the area of surface stabilization involve the treatment of the disturbed land surface and the problems associated with the disposal and reclamation of the waste material. These technologies are outlined in Figure 5.3-11 and their key features are presented in Table 5.3-5.

Dust Control--

The purpose of dust suppression is to limit pollution from airborne dust, particularly during the placement of the waste material in a fill. Dust suppression can be accomplished by spraying the haul roads and fill surface with water or a combination of water and a chemical binder. Haul roads could, alternatively, be paved.

Use of water alone for dust suppression would necessitate repeated applications, often more than one per day, to be effective. Water with a chemical binder should necessitate only a few applications to a given surface to stabilize it for a year or more unless it receives heavy traffic. Finally, vegetation would provide perhaps the most permanent means of dust control, but this would not be practical except on surfaces which would not be disturbed for a number of years.

The dust suppression technology assumed in developing the cost data for two examples consisted of routine spraying of the processed shale pile with water and additives to minimize the dust generated due to the wind and the



NOTES:

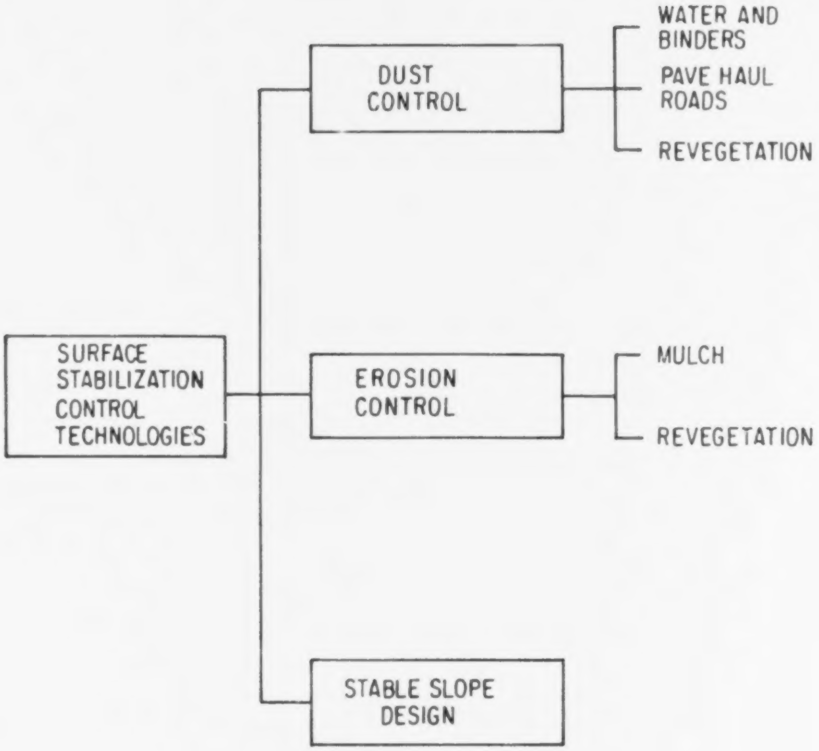
Examples 1 & 2 consist of gravel blankets for collection of groundwater from springs and seeps; extent of blankets dictated by the existence and extent of such conditions.

The costs indicated are cumulative for the project life.

See Section 6.2.3 for details on the solid waste management cost methodology.

SOURCE: SWEC

FIGURE 5.3-10 GROUNDWATER COLLECTION COSTS


SURFACE STABILIZATION CONTROL TECHNOLOGIES
DUST CONTROL
WATER AND BINDERS
PAVE HAUL ROADS
REVEGETATION
EROSION CONTROL
MULCH
REVEGETATION
STABLE SLOPE DESIGN

SOURCE: SWEC

FIGURE 5.3-11 SURFACE STABILIZATION TECHNOLOGIES

TABLE 5.3-5. KEY FEATURES OF SURFACE STABILIZATION TECHNOLOGIES

Control Technology	Principle	Purpose	Comments
Dust Control		Prevents or limits dust pollution from wind blowing across exposed surfaces or from vehicular traffic.	
Water and Binders	Fluid sprayed on the surface binds the fine particles together.		Well developed technology that is commonly used in mining operations.
Pave Haul Roads	A hard surface on the haul road prevents generation of dust by vehicular traffic.		Should improve traffic conditions on the road.
Revegetation	Vegetation prevents dust caused by wind.		Not useful in areas with any equipment traffic or where the surface is being disturbed by other activities.
Erosion Control		Simplifies reclamation, prevents blockage of the drains, and prevents contamination of surface waters by eroded material.	
Mulch	Various materials are placed on the slope to limit erosion.		Quick and easy to accomplish but is only a temporary measure.
Revegetation	Plant growth is started on the slope to limit erosion.		Permanent control but slow to achieve.
Stable Slope Design	Design slope to minimize stability problems and maintenance.	Makes erosion control, revegetation, and drainage easier. Restricts waste material to a definite, predefined area.	

Source: SWEC.

waste hauling and placement activities. Depending on the processed shale characteristics, this operation could either be continuous or intermittent. The cost curve in Figure 5.3-12 is based on the assumption that both the manpower and equipment operation requirements are continuous. Theoretically, these requirements could differ depending on the rate of waste production and the surface area of the particular waste pile; however, both cases estimated were assumed to be equivalent in this respect.

Erosion Control--

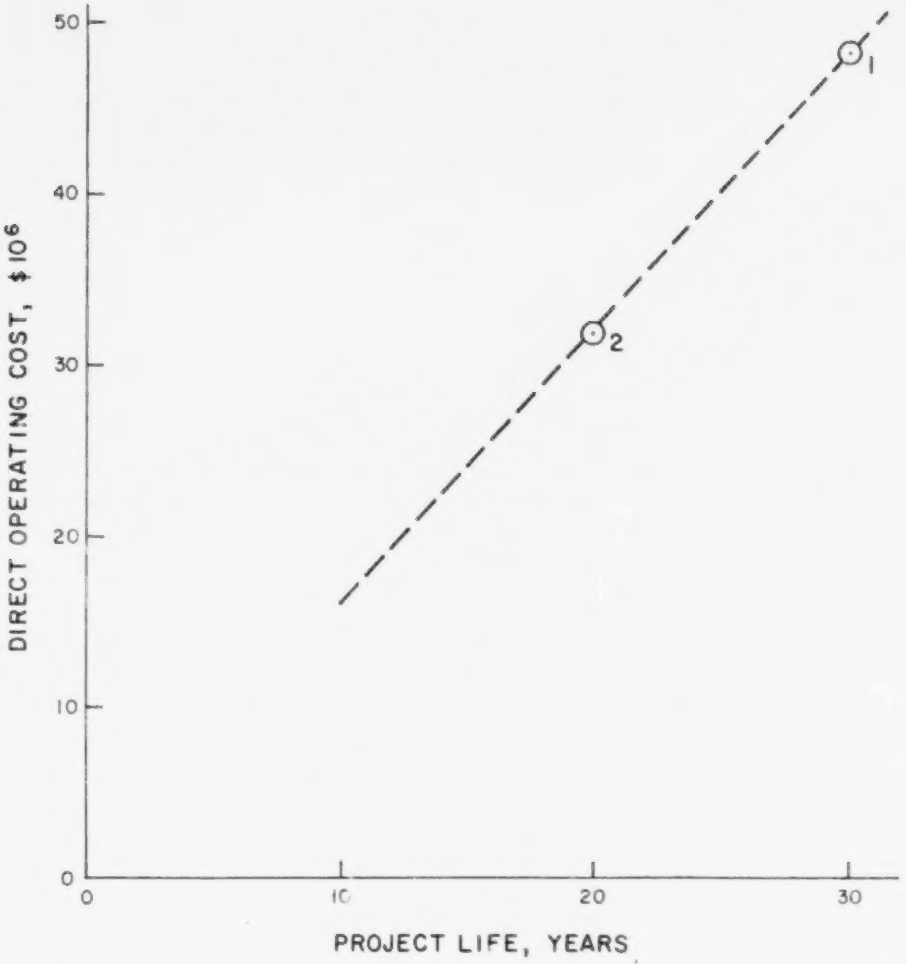
The purpose of erosion control is to keep the waste material in place so that the surface drains remain free flowing, the slopes remain stable, eroded material does not pollute surface streams, and reclamation and revegetation efforts are not hampered. Some means of limiting erosion include contouring the surface with short and gentle slopes, providing for drainage of the slopes at frequent intervals, using mulch or filter fabric to dampen the impact of water flow, and revegetating the completed faces. Of these measures, grading and drainage are essential, take effect immediately, and last as long as they are maintained. Mulch or filter fabric also provide a quick control, but they are of a temporary nature. Revegetation provides a permanent control, but it is generally slower to take effect.

A major consideration in planning erosion control measures is the severity of rainfall in the area. A large proportion of the water from a high-intensity rainfall would run off the surface, thus increasing the erosion.

Reclamation and revegetation consist of placing a subsoil and topsoil strata of sufficient thickness to support vegetation, and then seeding the disposal area with native or introduced species. The greatest contributor to the magnitude of cost for this control technology is the thickness of the soil strata and the costs associated with the delivered soil material, i.e., the source development, processing and hauling costs. Soil and subsoil stripped from the disposal site may not be available in sufficient quantity to meet the reclamation needs. The cost curves presented in Figure 5.3-13 illustrate five examples. Examples 1 and 5 included 2 feet of subsoil (sand-gravel material) and 30 inches of topsoil, both of which were brought in from off-site sources and thus had additional costs involved. Examples 2 and 3 also used the same thicknesses, but the soils were available on the site. Example 4 used no subsoil and only 6 inches of topsoil which was available on the site; therefore, additional material costs were not involved. All examples included the cost of revegetation. It is evident from the figure that the cost of erosion control can vary significantly depending on the factors considered; however, in any category, the costs are proportional to the area reclaimed and revegetated.

Stable Slope Design--

The purpose of designing the slopes to be stable under prevailing conditions is to minimize the maintenance of the landfill and to avoid hampering of the reclamation and revegetation efforts. The techniques used in designing stable slopes are a well developed part of soils engineering.



NOTES:

Example 1 assumes a 30-year project life, while Example 2 assumes a 20-year life.

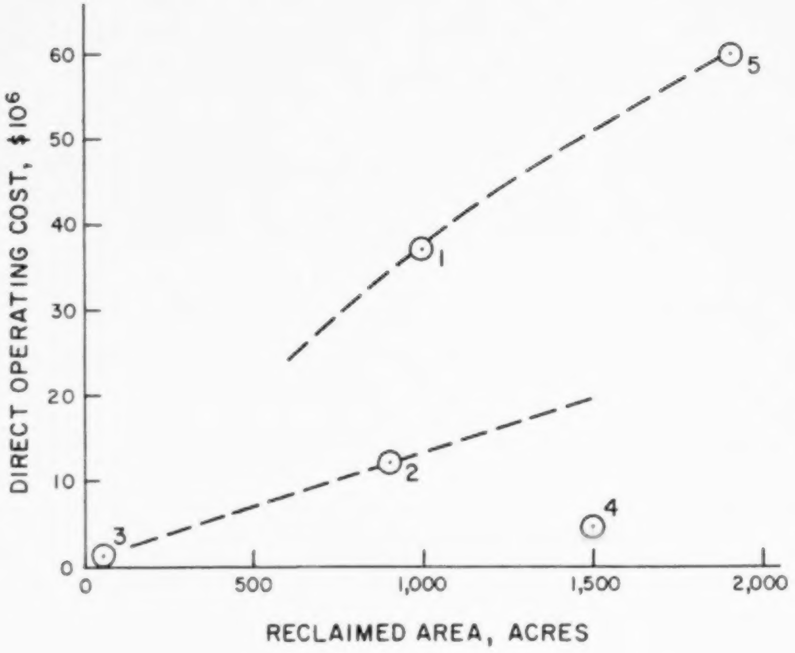
The Lurgi-Open Pit Case Study has a project life of 20 years.

The costs indicated are cumulative for the project life.

See Section 6.2.3 for details on the solid waste management cost methodology.

SOURCE: SWEC

FIGURE 5.3-12 DUST CONTROL COSTS



NOTES:

Examples 1 & 5 include 2 feet of subsoil and 30 inches of topsoil, both obtained off site.

Examples 2 & 3 include 2 feet of subsoil and 30 inches of topsoil obtained on site.

Example 4 includes no subsoil and only 6 inches of topsoil obtained on site.

The costs indicated are cumulative for the project life.

See Section 6.2.3 for details on the solid waste management cost methodology.

SOURCE: SWEC

FIGURE 5.3-13 RECLAMATION AND REVEGETATION COSTS

To arrive at the most advantageous slope design, other factors besides basic stability, such as erosion, ease of placement, reclamation and revegetation, must be considered. However, the physical characteristics of the waste material will dictate a limiting slope angle. The costs of achieving a stable slope design are incidental to the placement and revegetation of the fill material; hence, additional costs are not involved.

5.3.5 Hazardous Waste Control Technologies

The control of hazardous waste involves its permanent impoundment in a permitted disposal facility. This facility may be built on the project site or the wastes may be sent to an existent, off-site permitted facility. These options are outlined in Figure 5.3-14 and their key features are presented in Table 5.3-6.

On-site Disposal--

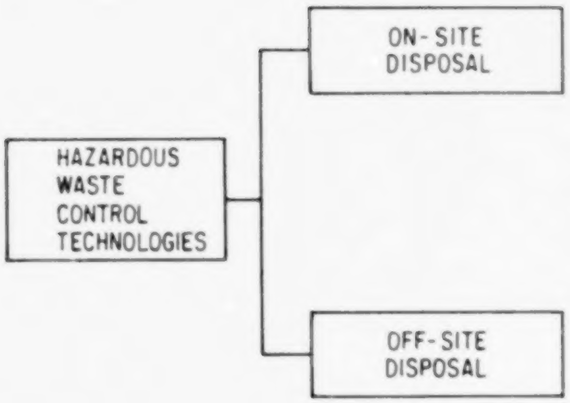
Hazardous waste lagoons are a well developed and accepted approach to solid waste management. They are actually an integration of several control technologies discussed in Sections 5.3.2, 5.3.3 and 5.3.4. Some of the included technologies would be an embankment surrounding the lagoon, a runon diversion system, one or two bottom liners, a surface cover, reclamation and revegetation, and monitoring.

There are certain advantages to building a hazardous waste facility on site. This option automatically assumes segregation of the hazardous and nonhazardous wastes and, hence, their separate disposal. An advantage of this approach is that much of the material necessary for the lagoon would be available on site or it already would have been brought in for the nonhazardous waste landfill. Furthermore, transport of the wastes beyond the property boundaries will not be required. A significant advantage may be that the producer of the hazardous wastes (the oil shale developer) will have complete control over the disposal of the wastes.

There are also certain disadvantages to on-site disposal of the hazardous wastes. To be efficient in evaporating the liquids and consolidating the sludge, the lagoon should be located preferably on a level site, which may not be readily available. Rugged, uneven terrain would increase the cost of site preparation, runon control and reclamation. There is also a possibility that the lagoon may interfere with other ongoing activities and the resource recovery.

Off-site Disposal--

Off-site existent facility. This would be an already existing facility where the wastes can be disposed of on an "as needed" basis. A payment is required for every shipment, but the cost may be lower than that of building and maintaining a new facility. Also, a significant amount of time and effort involved in the licensing, design, and construction of a new facility can be saved. The capacity and distance of the existent facility must also be considered in selecting the disposal approach.


HAZARDOUS WASTE CONTROL TECHNOLOGIES
ON-SITE DISPOSAL
OFF-SITE DISPOSAL

SOURCE: SWEC

FIGURE 5.3-14 HAZARDOUS WASTE CONTROL TECHNOLOGIES

TABLE 5.3-6. KEY FEATURES OF HAZARDOUS WASTE CONTROL TECHNOLOGIES

Control Technology	Principle	Purpose	Comments
On-site Disposal	Dispose of hazardous wastes in a lagoon established on site.	Dispose of hazardous wastes produced by processing of oil shales.	The oil shale developer has complete control of hazardous wastes produced by the facility.
Off-site Disposal	Establish lagoon off site or pay for disposal in existing permitted facility.	Dispose of hazardous wastes produced by processing of oil shales.	Provides a broader selection of sites, although the wastes must be transported. Potentially less involvement with the wastes.

Source: SWEC.

SECTION 6

POLLUTION CONTROL COSTS

This section provides an analysis of estimated pollution control costs for the Lurgi-Open Pit case study analyzed in this manual (see Sections 2 and 3 for a description of the case study). Section 6.1 presents fixed capital and direct annual operating costs for each control and explains how they were developed. These costs are referred to as the "engineering costs."

Section 6.2 explains the cost analysis methodology used to develop the total annual and per-barrel pollution control costs. These costs combine capital and annual operating costs, allow for taxes, and incorporate a return on investment. This is an approach similar to that which a private developer might use to determine costs or assess the economic feasibility of a project. Section 6.2 also details the economic assumptions that are incorporated into the calculation of total annual control costs.

Section 6.3 presents estimated total annual control costs and per-barrel costs for each control using a set of standard economic assumptions. These costs are assembled into total per-barrel costs for air and water pollution control for the case study examined in this manual. This section also examines the sensitivity of the per-barrel control costs to a series of changes in the engineering costs and economic assumptions.

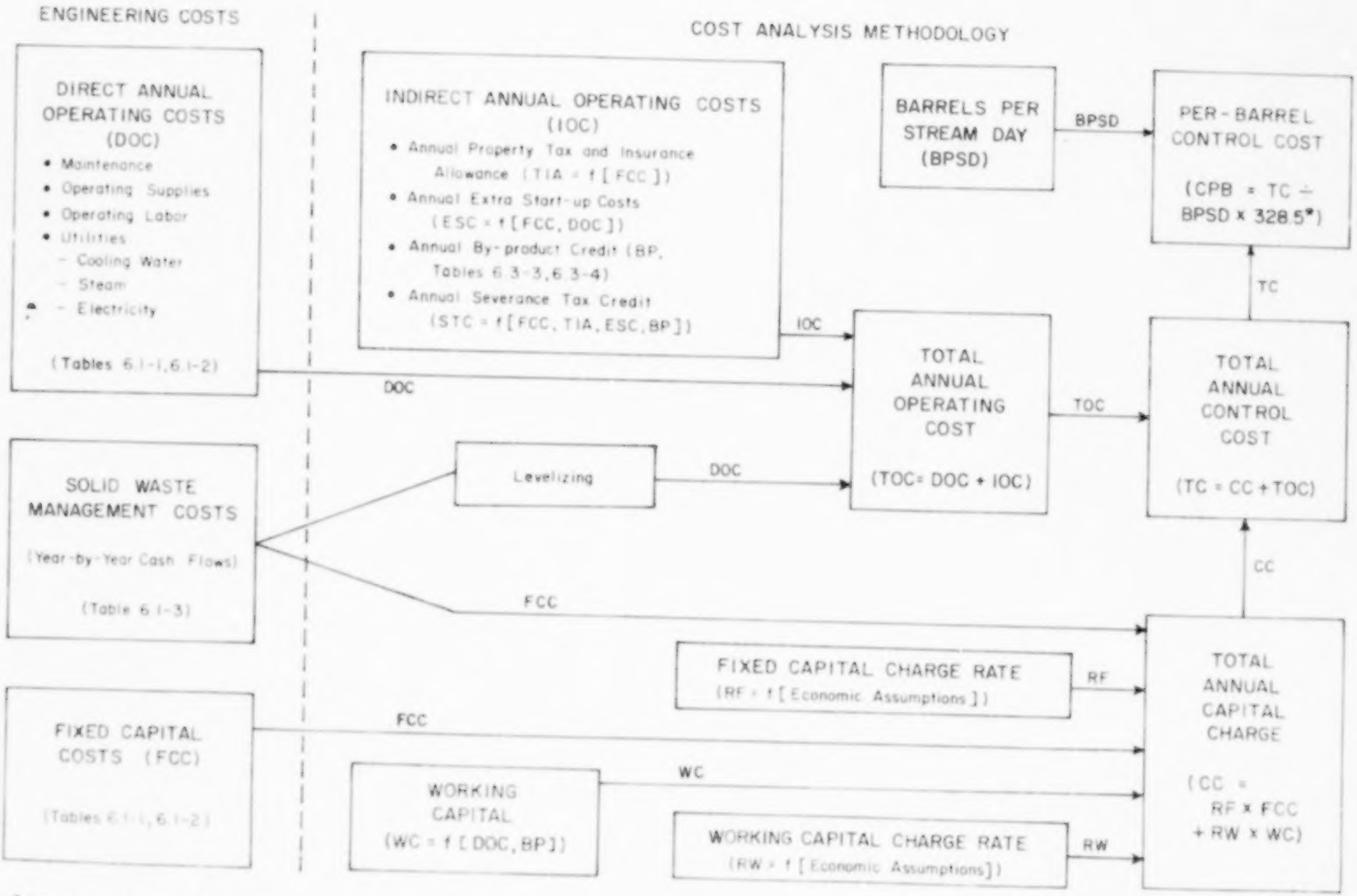
Section 6.4 provides more detailed information supporting Sections 6.1, 6.2 and 6.3. Section 6.4.1 provides the algorithms that were used to determine total annual control costs and per-barrel control costs, and Sections 6.4.2 and 6.4.3 provide examples, respectively, of fixed charge rate calculations and cost levelizing calculations.

Section 6 uses a large number of cost and economic terms. The interrelationships among the more important of these terms is illustrated in Figure 6.0-1. Each term is explained when it is first used in the text, but the reader may find it helpful to use this figure to provide an overview while reading the various sections. In addition, Table 6.2-4, presented later in this section, indicates the estimated relative magnitude of the components of per-barrel control cost for a typical major pollution control.

6.1 ENGINEERING COST DATA

6.1.1 Bases of Engineering Cost Data

Throughout this manual a distinction is made between capital costs and annual operating costs. There are two types of capital cost, fixed capital


ENGINEERING COSTS
COST ANALYSIS METHODOLOGY
DIRECT ANNUAL OPERATING COSTS (DOC)
• Maintenance
• Operating Supplies
• Operating Labor
• Utilities
– Cooling Water
– Steam
– Electricity
(Tables 6.1-1, 6.1-2)
INDIRECT ANNUAL OPERATING COSTS (IOC)
• Annual Property Tax and Insurance Allowance (TIA = f [FCC])
• Annual Extra Start-up Costs (ESC = f [FCC, DOC])
• Annual By-product Credit (BP, Tables 6.3-3, 6.3-4)
• Annual Severance Tax Credit (STC = f [FCC, TIA, ESC, BP])
BARRELS PER STREAM DAY (BPSD)
BPSD
PER-BARREL CONTROL COST
(CPB = TC ÷ BPSD x 328.5*)
TC
IOC
DOC
TOTAL ANNUAL OPERATING COST
(TOC= DOC + IOC)
TOC
TOTAL ANNUAL CONTROL COST
(TC = CC + TOC)
SOLID WASTE MANAGEMENT COSTS
(Year-by-Year Cash Flows)
(Table 6.1-3)
Levelizing
DOC
FCC
CC
FIXED CAPITAL CHARGE RATE
(RF = f [Economic Assumptions])
RF
TOTAL ANNUAL CAPITAL CHARGE
(CC = RF x FCC + RW x WC)
FIXED CAPITAL COSTS (FCC)
(Tables 6.1-1, 6.1-2)
FCC
WORKING CAPITAL
(WC = f [DOC, BP])
WC
WORKING CAPITAL CHARGE RATE
(RW = f [Economic Assumptions])
RW

*328.5 is the number of operating days in a normal year

Note: f means "a function of"

SOURCE: DRI

FIGURE 6.0-1 INTERRELATIONSHIPS AMONG VARIOUS COST AND ECONOMIC TERMS

and working capital, and two types of annual operating cost, direct and indirect.

Fixed capital is investment in construction and equipment, whereas working capital is money that is required to operate the plant, e.g., that which is tied up in inventories.

Direct annual operating costs include maintenance, operating supplies, operating labor and utilities costs. Indirect annual operating costs comprise additional annual costs, i.e., property tax and insurance, an allowance for extra start-up costs, a credit for severance tax not paid and by-product credits.

Section 6.1 only considers fixed capital costs and direct annual operating costs. Working capital and indirect annual operating costs are considered in Section 6.2.

Assumptions Used to Develop Costs--

All costs are expressed in mid-1980 constant dollars. The following data apply to air and water pollution control costs. Solid waste management costs were developed on the basis that these activities are contracted out, since they are all construction-type activities (see discussion later in this subsection).

Fixed capital costs. Fixed capital costs are of the "preliminary estimate" category. Physical plant costs for air emission controls were developed by Stone and Webster Engineering Corporation (SWEC) and for water pollution controls by Water Purification Associates (WPA). Actual vendor quotes were used for major items of equipment; costs for other equipment were obtained from data files maintained by SWEC and WPA. Total physical plant costs were developed from the equipment costs by adding appropriate allowances for the following:

- Site preparation, excavation and foundations
- Concrete and rebar
- Support structures
- Piping, ductwork, joints, valves, dampers, etc.
- Duct and pipe insulation
- Pumps and blowers
- Electrical
- Instrumentation and controls
- Monitoring equipment
- Erection and commissioning
- Painting
- Buildings.

To arrive at the total fixed capital cost, the following factors were added to the physical plant cost:

Engineering and construction overhead:	25% of physical plant cost.
Contractor's fee:	3% of bare module cost (physical plant cost plus engineering and construction overhead).
Contingency:	20% of bare module cost.

For an explanation of this method of developing estimates of fixed capital costs, see Uhl (June 1979). A 20% contingency factor was chosen because there are only pilot plant data for the Lurgi retorting process.

It is considered that the accuracy of these cost estimates is within ±30 percent. Although the accuracy of a preliminary fixed capital cost estimate is normally regarded as ±20 percent, uncertainties about stream magnitudes and composition decrease the accuracy of these estimates to ±30 percent.

Direct annual operating costs. There are two components which make up the total annual operating cost. The direct annual operating cost can be regarded as the basic (or engineering) cost, while calculation of the indirect annual operating cost makes some adjustments to this cost. By-product credits are included in the indirect annual operating cost. Data on the bases of direct annual operating costs are given below, while the bases of indirect annual operating costs are outlined in Section 6.2.

Direct annual operating costs are made up of the following components:

- Maintenance
- Operating supplies
- Operating labor
- Utilities
  - --Cooling water
  - --Steam
  - --Electricity.

Maintenance costs include maintenance labor and replacement parts, consumables used for maintenance, etc.

Operating supplies are consumable items (such as chemicals) used in the regular operation of the control (as opposed to use for maintenance).

Operating (and maintenance) labor is costed at \$30/hr. This is a "loaded" rate, meaning that it incorporates some overhead-type costs to avoid developing them separately. The rate is made up as follows:

A.	Wages for direct labor	\$11.00/hr
B.	Fringe benefits (45% of A)	4.95
C.	Field supervision (15% of A + B)	2.40
D.	Overhead (50% of A + B + C)	9.20
E.	General & administrative charge (9% of A + B + C + D)	2.45
	Total	\$30.00/hr

In mid-1980, examination of union agreements showed that oil refinery direct operating labor was receiving approximately \$10/hr in Colorado. However, it is anticipated that when oil shale development occurs, this will bid up local labor rates, so \$11/hr, which was used for the oil shale PCTMs, is a reasonable value. The multiplier factors, used to arrive at the "loaded" labor rate of \$30/hr, were suggested by SWEC based on project experience in the western U.S.A.

Cooling water is costed at 11.3 cents per 10^3 gal circulated (3¢/m^3). This is only a charge for the use of the cooling tower. The cost of treating the makeup water is included under water pollution control.

Process steam is charged at \$3.00 per million Btu.

Electricity is charged at 3 cents per kW-hr.*

There is no contingency factor in the direct annual operating costs for air and water pollution controls.

Solid Waste Management Costs--

Solid waste management costs in the form of year-by-year cash flows were developed by SWEC using company cost data files. They include the same engineering and construction overhead, contractor's fee, and contingency factor (20%) as the fixed capital costs discussed earlier. The use of a 20% contingency factor is appropriate since all solid waste management costs are of a construction nature, subject to uncertainties similar to those inherent in fixed capital costs.

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* To be consistent among the three oil shale PCTMs, electricity is charged at 3 cents per kW-hr, whether purchased or generated on site. This figure represents a compromise between the value of electricity sold by plants that will have surplus on-site generated power and the higher cost of power purchased from a utility.

6.1.2 Details of Engineering Costs

Tables 6.1-1 and 6.1-2 present details of the fixed capital and direct annual operating costs for each air and water pollution control. The operating costs relate to a year of normal operation, i.e., full production. For the start-up period, direct annual operating costs are modified to an appropriate level by the cost analysis methodology.

Table 6.1-3 details the solid waste management costs on a year-by-year basis. These costs are allocated to fixed capital or direct annual operating categories in Section 6.2 (Table 6.2-3). Insufficient information was available to develop a complete plan for solid waste management operations. Consequently, the solid waste management costs presented here are for certain items only and do not represent the total pollution control cost for solid waste.

6.2 COST ANALYSIS METHODOLOGY

In the cost analysis, engineering cost data are transformed into two primary measures--the total annual pollution control cost and the control cost per barrel of shale oil. These costs incorporate both capital and annual operating costs and consider project timing, taxes, and the necessary return on investment.

6.2.1 Overview of Cost Analysis Methodology

In private industry, one of the most widely accepted methods of evaluating the economics of a project is the discounted cash flow (DCF) approach. Using this approach, a project must be able to demonstrate that it can produce some established minimum rate of return on investment--known as a "hurdle" rate--to be acceptable.

One method for applying the DCF approach to a complete oil shale project is to determine the selling price which would provide the revenue required to produce a minimum acceptable rate of return (DCF ROR). With this method, a selling price for oil can be established by distributing the required revenue uniformly over every barrel of oil produced.

The same technique can be utilized to determine the total annual and per-barrel costs of pollution control. In practice, pollution control is not a separable aspect of an oil shale project. Consequently, a private developer will require the same DCF ROR on pollution controls as for the entire project.

If the revenue necessary to provide the required DCF ROR for each control (expressed in constant dollars) is distributed uniformly over each barrel of shale oil produced, then this also implies a constant total revenue requirement in each year of normal (full) production. However, in the start-up years, less oil is produced, with the result that the annual revenue requirement is prorated. Additional costs incurred in the start-up period were spread over all production in order to produce a uniform per-barrel control cost.

TABLE 6.1-1. DETAILED ENGINEERING COSTS FOR AIR POLLUTION CONTROLS

Control (No. of Units)	Control Location	Fixed Capital Cost ($000's)	Components of Direct Annual Operating Cost ($000's/yr)				Total Direct Annual Operating Cost ($000's/yr)
			Maintenance	Operating Supplies	Operating Labor	Electricity	
Particulate Controls							
Fabric Filters (2)	Primary Crusher (ore)	987	19	--	--	44	63
Fabric Filter (1)	Primary Crusher (subore)	105	2	--	--	4	6
Fabric Filter (1)	Primary Crusher (overburden)	552	11	--	--	23	34
Fabric Filters (3)	Raw Shale Conveyor Transfer Points	1,051	21	--	--	44	65
Fabric Filters (2)	Conveyor to Stockpile	628	12	--	--	73	85
Fabric Filters (8)	Secondary Crushers	4,828	94	--	--	203	297
Fabric Filters (8)	Secondary Screens	4,828	94	--	--	203	297
Fabric Filters (9)	Tertiary Crushers	5,432	106	--	--	228	334
Fabric Filters (9)	Tertiary Screens	5,432	106	--	--	228	334
Fabric Filter (1)	Fine Ore Storage	249	5	--	--	10	15
Fabric Filters (2)	Processed Shale Conveyor Transfer Points	559	11	--	--	23	34
Fabric Filters (3)	Processed Shale Load-out Hoppers	559	11	--	--	23	34
Fabric Filters (2)	Conveyor to Secondary Crushers	348	7	--	--	15	22
Fabric Filters (2)	Conveyor to Secondary Screens	348	7	--	--	15	22
Fabric Filters (2)	Conveyor to Tertiary Crushers	348	7	--	--	15	22
Fabric Filters (4)	Conveyor to Tertiary Screens	696	14	--	--	30	44
Fabric Filters (2)	Conveyor to Fine Ore Storage	348	7	--	--	15	22
Fabric Filters (2)	Conveyor to Retort Feed Hoppers	348	7	--	--	15	22
Water and Foam Sprays	Open Stockpiles, etc.	909	117	1,065	--	274	1,456
Fabric Filters (4)	Retort Feed Hoppers	1,837	36	--	--	77	113
Fabric Filters (13)	Conveyor to Retorts	2,059	40	--	--	87	127
Flue Gas Treatment							
Electrostatic Precipitators (13)	Flue Gas Discharge System	50,734	330	--	--	1,814	2,144

(Continued)

TABLE 6.1-1 (cont.)

Control (No. of Units)	Control Location	Fixed Capital Cost ($000's)	Components of Direct Annual Operating Cost ($000's/yr)				Total Direct Annual Operating Cost ($000's/yr)
			Maintenance	Operating Supplies	Operating Labor	Electricity	
Miscellaneous Controls							
Stretford (1)	DEA Unit	6,860	134	164	350	121*	769
Ammonia Storage Tank (1)	Ammonia Recovery	466	--	--	--	--	--
Floating Roof Oil Storage Tanks (2)	Product Storage	300	--	--	--	--	--
Proper Maintenance	Valves, Pumps, etc.	61	55	--	6	--	61
Catalytic Converters	Diesel Equipment	170	60	--	--	--	60

* This includes $24,000 for steam.

Source: DRI estimates based on information provided by SWEC.

TABLE 6.1-2. DETAILED ENGINEERING COSTS FOR WATER POLLUTION CONTROLS

Control	Fixed Capital Cost ($000's)	Components of Direct Annual Operating Cost ($000's/yr)						Total Direct Annual Operating Cost ($000's/yr)
		Maintenance	Operating Supplies	Operating Labor	Cooling Water	Steam	Electricity	
Ammonia Recovery Unit	3,627	118	428	237	60	1,565	11	2,419
API Oil/Water Separator	161	4	--	--	--	--	--	4
Mine Water Clarifier*	2,560	84	235	--	--	--	--	319
Cooling Water Treatment*	--	--	51	--	--	--	--	51
Boiler Feedwater Treatment*	122	4	14	40	--	--	11	69
Equalization Pond	181	3	--	--	--	--	--	3
Runoff Oil/Water Separator	41	1	--	--	--	--	--	1
Aeration Pond	430	14	--	99	--	--	40	153
TOTAL	7,122	228	728	376	60	1,565	62	3,019

* These technologies could be considered as part of the process rather than pollution control.

Source: DRI estimates based on information provided by WPA.

TABLE 6.1-3. **ENGINEERING COSTS AND TIMING OF SOLID WASTE MANAGEMENT ACTIVITIES**
(Thousands of Dollars)

Activity	Project Year → 1	2	3	4	5	6	7	8	9	10
SURFACE HYDROLOGY										
Runoff Collection Sumps	58	58	58	58	58	58	58	58	58	58
Runoff Collection Pumps	15	15	15	15	15	15	15	15	15	15
Deep Monitoring Wells	858									
Shallow Monitoring Wells										26
Piezometers										432
SURFACE STABILIZATION										
Dust Suppression	6,204	9,196	11,079	11,079	11,079	11,079	11,079	11,079	11,079	11,079
Revegetation										
Topsoil										
Seed										

Activity	Project Year → 11	12	13	14	15	16	17	18	19	20	21
SURFACE HYDROLOGY											
Runoff Collection Sumps	58	58	58	58	58	58	58	58	58	58	
Runoff Collection Pumps	15	15	15	15	15	15	15	15	15	15	
Deep Monitoring Wells											
Shallow Monitoring Wells											
Piezometers											
SURFACE STABILIZATION											
Dust Suppression	11,079	11,079	11,079	11,079	11,079	11,079	11,079	11,079	11,079	11,079	
Revegetation									185	185	185
Topsoil									46	46	46
Seed									11	12	11

Note: Year 1 is the first year of production. This is subsequent to the 30-year open pit development period.

Source: DRI estimates based on information provided by SWEC.

The total annual required revenue is utilized to satisfy two major components: the total annual operating cost, and a component that provides the necessary return on investment, called the total annual capital charge. Note that with the DCF approach, profit is based solely on investment; operating costs are passed straight through as one component of the total revenue requirement, without addition of any profit element. This is normal practice for industrial project assessments.

To relate an annual capital charge to the corresponding investment, a "capital charge rate" was used. In practice, there are two types of capital investment: fixed capital (i.e., physical equipment) and working capital (which is nondepreciable investment). The "fixed charge rate" is defined as the proportion of investment in fixed capital that must be recovered in a year of normal production in order to provide the required DCF ROR. The "working capital charge rate" performs a similar function for the working capital. The total annual capital charge for a pollution control is the sum of the annual fixed capital charge and the annual working capital charge.

Fixed charge rates have several economic assumptions embedded in them. Some of these assumptions are common to all pollution controls, i.e., the project life and operating (stream) factors, the income tax rate, and the required DCF ROR.

Other assumptions vary according to the pollution control or group of controls. These are: the timing of the investment in fixed capital, the depreciation period, and the investment tax credit details. Consequently, different fixed charge rates are used for different groups of pollution controls.* (These rates, as well as the underlying standard economic assumptions, are listed later in Table 6.2-2.)

The working capital charge rate depends only on the project life and operating factors, the timing of the investment in working capital and the required DCF ROR. Since none of these assumptions varies among controls, the same working capital charge rate is used for each control.

As already indicated, the total annual cost for a control is the sum of the total annual capital charge and the total annual operating cost. The total annual operating cost comprises two components. The "direct annual operating cost" consists of maintenance, operating supplies, operating labor and utilities. The "indirect annual operating cost" comprises an annual allowance for property taxes and insurance, any annual by-product credits, and an allowance for extra start-up costs, i.e., those that are in excess of the direct annual operating cost prorated in accordance with production. It also includes a credit reflecting a reduction in the Colorado severance tax that must be paid, because the cost of each pollution control reduces the severance tax liability.* Extra start-up costs and the severance tax credit are "levelized" to distribute them uniformly over each barrel of shale oil produced since they do not vary in proportion to production. (Levelizing takes a cost that does not vary in proportion to production and finds an economically equivalent cost that has the same time-profile as production [see Sections 6.2.3 and 6.4.3].) To summarize:

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* The use of several different fixed charge rates in the same oil shale PCTM may appear complex. However, since the manuals examine several alternatives for pollution control, an accurate evaluation of capital charges is needed. A less accurate approach, such as assuming a single capital expenditure profile for all controls, could conceivably affect the per-barrel cost ranking of pollution control alternatives.

> Total Annual Control Cost = Annual Fixed Capital Charge + Annual Working Capital Charge + Direct Annual Operating Cost + Indirect Annual Operating Cost.

For air and water pollution controls, direct annual operating costs are specified for a normal year of production and are implicitly prorated during the start-up years. In practice, operating costs during the start-up period will be higher, but this is allowed for via the extra start-up costs discussed in Section 6.2.2. The solid waste management costs are developed in the form of a year-by-year cash flow (see Table 6.1-3) which must be converted into equivalent fixed capital and direct annual operating costs for a full production year (see Section 6.2.3 and Table 6.2-3).

The per-barrel control cost is obtained by dividing the total annual control cost by the production in a normal (full production) year. (Per-barrel operating costs and capital charges can be calculated in the same way.) The detailed algorithms for these calculations and for determining fixed and working capital charge factors are given in Section 6.4.1.

6.2.2 Economic Assumptions Used in Total Cost Calculations

To transform engineering cost data provided in Section 6.1.2 into total annual capital charges, total annual operating costs, and total annual or per-barrel control costs, a number of economic assumptions were made. Most of these assumptions are listed in Table 6.2-1, and Table 6.2-2 summarizes those assumptions that vary from control to control. The values given in these two tables are the standard values, known as the "standard economic assumptions," which have been used for the cost analyses presented in the oil shale PCTMs. Some of these are varied in the sensitivity analyses which are used to show how control costs change in response to alternative economic assumptions and to changes in the engineering costs.

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* The distinction between the two components of operating cost is made for convenience in performing the calculations and is not fundamental. The direct annual operating cost is comprised of basic cost elements, whereas the indirect annual operating cost comprises a series of adjustments that are influenced by other factors, such as tax assumptions. Direct annual operating costs for each control are given in Tables 6.1-1, 6.1-2 and 6.2-3. Indirect annual operating costs for all controls are calculated using a standard algorithm (see Section 6.2.2), except for any by-product credits which are given in Tables 6.3.3 and 6.3.4.

TABLE 6.2-1. SUMMARY OF STANDARD COST AND ECONOMIC ASSUMPTIONS

Assumptions

COST ASSUMPTIONS

- Base Year: Mid-1980 dollars
- Direct Labor Rate: $11.00/hr*
- "Loaded" Labor Rate*: $30.00/hr
- Fixed Capital Costs: 25% engineering and construction overhead and 3% contractor's fee included*
- Contingency Allowances: 20%, all fixed capital costs*
  0%, most operating costs*
  20%, solid waste direct operating costs

ECONOMIC ASSUMPTIONS

- Project Life: 20 years*
- Normal Output: 63,140 Barrels per Stream Day (BPSD)
- Operating (stream) Factors: Year 1 - 50%
  Year 2 - 75%
  Years 3-20 - 90%*
- Approach: Discounted Cash Flow Evaluation (DCF)*
- Discount Factors: Discrete,* year-end basis
- Method: Determination of Revenue Required to provide specified DCF ROR*
- Technique: Annual Capital Charge plus Annual Operating Cost
- Required DCF ROR: 12% (100% Equity Basis)*
- Cost Escalation: None (constant dollar evaluation)*
- Combined State and Federal Income Tax Rate: 48%*
- Depreciation: Method - Sum-of-Year's Digits*
  Period - 16 years, most items*
  10 years, solid waste area
  5 years, mobile equipment
- Investment Tax Credit: 20%, most items*
  13 1/3%, mobile equipment
- Additional Start-up Costs (in Year 1): 3% of fixed capital, plus 20% of a normal year's direct operating cost
- Working Capital: 30 days' total operating cost (excluding by-product credit), plus 60 days' by-product credit
- Annual Allowance for Property Taxes and Insurance: 3% of fixed capital
- Colorado Severance Tax: Credit allowed
- Timing of Investment: Initial fixed capital expenditures can occur in Years -3 through +1; expenditures and tax considerations for each control are phased in accordance with the construction and initial operation of each control (see Table 6.2-2 for schedules)
- Corporate Financing: Tax credits and allowances can be passed through to a parent company that can benefit from them immediately, without waiting for the project to become profitable*
- Federal Depletion Allowance: Does not affect pollution control costs

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* These methods and factors are in accordance with the recommendations, dated April 22, 1980, of EPA's ad hoc synfuels cost committee.

Source: DRI.

TABLE 6.2-2. ECONOMIC ASSUMPTIONS THAT VARY FROM CONTROL TO CONTROL

	Capital Expenditure	Investment Tax Credit		Depreciation		Fixed Charge Rate[a]
	Profile	Rate %	Profile	Life (years)	Starts	Percent
Retort Timing Controls associated with retorting: certain fabric filters, electrostatic precipitators, Stretford, ammonia recovery unit, API oil/water separator, boiler feedwater and cooling water treatment	Year -2: 10% Year -1: 30% Year 0: 60%	20	Same as capital[b]	16	Year +1	16.17
Mine Timing Controls governed by mine and project start up: most fabric filters, water and foam sprays, ammonia and oil storage, maintenance of valves, pumps, etc.	Year -1: 30% Year 0: 70%	20	Same as capital[b]	16	Year +1	15.61
Early Water Management Controls associated with mine and site water treatment: mine water clarifier, equalization pond, runoff oil/water separator, aeration pond	Year -3: 100%	20	Year -2: 100%	16	Year -2	21.64
Catalytic Converters[c] (on diesel equipment)	Year 0: 100% Year +7: 100% Year +14: 100%	13 1/3[d]	Year +1: 100% Year +8: 100% Year +15: 100%	5 5 5	Year +1: 100% Year +8: 100% Year +15: 100%	23.36
Solid Waste Management (Year 1) Deep monitoring wells	Year +1: 100%	20	Year +2: 100%	10	Year +2	12.49
Solid Waste Management (Year 10) Shallow monitoring wells and piezometers	Year +10: 100%	20	Year +11: 100%	10	Year +11	4.51

[a] For standard economic assumptions (see Table 6.2-1).

[b] Qualifies for investment tax credit progress payments.

[c] Capital is replaced twice during project life.

[d] Investment tax credit is reduced because equipment life is less than 7 years.

Source: DRI.

Where appropriate, the standard economic assumptions are discussed below. Others are discussed in connection with the sensitivity analyses in Section 6.3.2.

Timing of Control Capital Expenditures--

Table 6.2-2 includes the fixed capital expenditure profiles for each category of control. Although a number of developers and other organizations have published construction schedules for oil shale plants, no schedule is available that is appropriate to a Lurgi-Open Pit plant of this size. Instead, the schedule was based on data for a 51,500 BPSD TOSCO II plant for which comparatively good data are available (Nutter and Waitman, 1978; telephone interview with C. S. Waitman, Tosco Corp., February 1979; Colony Development Operation, 1977). Engineering judgment was then used to determine when the pollution controls would be procured and installed, incorporating the impact of payments made during off-site fabrication. In general, expenditures on pollution controls tend to be incurred later than those for most retort construction activities, since the controls are usually among the last items to be installed.

Part of the water pollution control system constitutes an exception to the above discussion. Basic site water management facilities must be installed and operational before most other activities can commence. Consequently, these items were assumed to be installed in Year -3 (i.e., 4 years before production commences) and placed into service in Year -2 for depreciation purposes. The mine water treatment system was given the same timing, but this is somewhat arbitrary since the mine is assumed to be fully developed at the commencement of this case study analysis. Also, because no mine development is included in this case study analysis, it was assumed that the mobile diesel equipment was purchased in Year 0 and placed into service in the first year of production, Year 1.

Assumptions for Taxation*--

Depreciation. All oil shale PCTMs used a 16-year depreciation period for most assets. This corresponds to the mid-point of the IRS' Asset Depreciation Range (ADR) guidelines for oil refineries. In practice, many companies would use the lower end of the ADR range, which is 13 years; however, it has been found that this would make very little difference in the results of the analysis.

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* All analyses were conducted prior to enactment of the Economic Recovery Tax Act of 1981 (PL 97-34). As far as an oil shale project is concerned, the main impact of this act is to permit very rapid depreciation under the Accelerated Cost Recovery System (ACRS). Using ACRS, most property would be depreciated over 5 years and mobile equipment would be depreciated over 3 years. A rough estimate of the effect of the provisions of the Economic Recovery Tax Act of 1981 on the pollution control costs is given in Section 6.3.1.

Some equipment clearly qualifies for a shorter life. Capital items associated with processed shale disposal, i.e., the monitoring wells and piezometers, were regarded as mining equipment, for which a 10-year depreciation period was used. A 5-year depreciation period was used for the mobile diesel equipment, and it was assumed that this equipment was replaced twice during the project life.

The depreciation method used for all taxation calculations was the Sum-of-the-Year's Digits method.

Investment Tax Credit (ITC). A basic 20% ITC was used for all items in accordance with the Energy Tax Act of 1978 (PL 95-618). The mobile equipment has a depreciation period of only 5 years, so the credit is reduced by one-third, to 13 1/3 percent.

Where payments for a control extend over more than one year, the tax credit can be taken as the capital is expended, in accordance with the IRS' progress payments rule. Otherwise, it is taken when the asset is placed into service.

Income tax rate. A combined State and Federal tax rate of 48% was used. In practice, Colorado has a 5% tax rate, so the effective percentage rate should be: 5 + ([1 - 0.05] x 46) = 48.7%. The error introduced by using 48% is negligible.

Depletion allowance. The Federal depletion allowance has not been incorporated into the calculation of taxes. The justification for this is as follows. The percentage depletion allowance is 15% on the "gross income" from an oil shale property. In this case, since the sales or transfer price of shale oil (and, hence, gross income) is independent of pollution control costs, the depletion allowance will not affect those costs. However, there is a limitation that the percentage depletion allowance cannot exceed 50% of the taxpayer's taxable income from the property, computed without allowance for depletion. Since pollution control costs reduce the taxable income, they could affect the depletion allowance if it was limited under the above rule, and this would then be a cost attributable to pollution control. While this might well be the case in a start-up year, it appears that this limit is unlikely to apply during a normal year's operation. This is because the complete project's total annual operating costs are a comparatively low proportion of its total annual costs, including capital-related costs (based on data for an open pit mine with unspecified type of surface retort producing 100,000 barrels per day [Peat, Marwick, Mitchell & Co., September 1980]).

Hence, the impact of the Federal percentage depletion allowance on pollution control costs has been disregarded. This may introduce minor errors during start-up years, but complete project cost data are not publicly available to permit the effect to be calculated. Cost depletion, which might at times be taken instead of percentage depletion, is clearly irrelevant to pollution control costs.

Other Assumptions--

DCF ROR. Twelve percent (per year) was used as a standard assumption (see Section 6.3.2).

Project life. The expected project life (measured from the commencement of production) will be determined by exhaustion of the oil shale reserves or by technological obsolescence. Planned project lives used for evaluations of oil shale developments range from 18 to 30 years. Twenty years is a common period to use for economic evaluations and was used in this manual. Increasing the life has a very small effect on the results at normal DCF RORs (i.e., 12% or more).

Start-up profile. The start-up profile and normal year operating factor are based on projections for a TOSCO II plant (Nutter and Waitman, 1978). Lurgi representatives consider that a Lurgi plant should achieve a better start-up profile than a TOSCO II plant, but they feel that a 90% operating factor may be slightly optimistic for a normal year (interview with H. Weiss and J. Arnhold of Lurgi Kohle und Mineralötechnik GmbH, in Denver, Colorado, January 1981). The operating (stream) factors used (i.e., Year 1: 50%, Year 2: 75%, Years 3-20: 90%) are considered to be the most appropriate assumptions that can be made at this time.

Components of Annual Indirect Operating Costs--

The annual indirect operating cost is composed as follows:

Annual property tax and insurance allowance

+ Extra start-up costs (levelized)

- Severance tax credit (levelized)

- Annual by-product credit (if any).

Property tax and insurance allowance. The annual indirect operating cost includes 3% of the fixed capital cost as an allowance for property tax and insurance. This value was selected by DRI after review of a wide variety of sources.

Extra start-up cost. The total extra start-up cost (which is treated as an operating cost, as opposed to being capitalized) is derived from the fixed capital and direct annual operating costs. The capital-related component is 3% of the fixed capital cost as an allowance for "fix it" costs. The operating cost-related component, which is 20% of a normal year's direct operating cost, allows for hiring and training employees before production commences and for higher unit costs during the start-up period. This value for the extra start-up cost for surface retorting plants with a 2-year start-up period was selected by DRI after a review of several sources, including estimates for TOSCO II (Nutter and Waitman, 1978) and Paraho (Pforzheimer and Kunchal, March 24, 1977) plants. The extra start-up cost was assumed to be incurred during the first year of production but is levelized to spread it uniformly over every barrel of oil produced (see Sections 6.4.1 and 6.4.3).

Severance tax credit. Under Colorado HB 1076, enacted in 1977, severance tax is levied on the production of a commercial oil shale facility at the rate of 4% of the "gross proceeds" for surface retorted oil. "Gross proceeds" is defined as the value of the oil shale at the point of severance and is calculated by subtracting costs (e.g., retorting and mining) from the gross sales income. Since pollution controls add to costs, they reduce the gross proceeds by a corresponding amount. Hence, a credit for severance tax not paid should be deducted from the pollution control costs.

While operating costs are clearly allowable in calculating gross proceeds, return on capital does not appear to be (the statute refers to allowing "...costs, including direct and indirect expenditures for: (a) equipment and machinery...."). Hence, when this credit is calculated, the capital charge must be replaced by some form of amortization. For this analysis, the severance tax credit calculations are based on direct and indirect annual operating costs, plus 5% of the fixed capital cost to provide capital amortization over the 20-year project life.

In applying this credit, allowance was also made for exemptions to the tax for the first 10,000 barrels per day of production and for plants that have not achieved 50% of their design capacity, together with reduced rates of tax in the early years. The credit is levelized in order to achieve a uniform per-barrel cost. The methodology utilized (LFAC2 in Section 6.4.1) is not precise, but since the severence tax correction is typically less than 2% of the total annual or per-barrel control cost (see Section 6.2.4), further refinement is not justified.*

By-product credits. The by-product credit (if any) for each control is shown in Tables 6.3-3 and 6.3-4. (There are no salable by-products from solid waste management.) By-product values of $110 per ton for ammonia, $30 per long ton for sulfur, and $32 per barrel for oils were used.

At present, there is no significant market for sulfur in the Rocky Mountain Region; in the past, shipping costs to move recovered sulfur to a chemical complex could have been greater than its delivered value. However, the price of high quality sulfur has gone up substantially in recent years, reaching values as high as $129 per long ton (U.S. DOI, August 1981). High demand for sulfur is projected through the year 2000 (Rangnow and Fasullo, September 28, 1981). Hence, a nominal $30 per long ton has been included for recovered sulfur. However, if in the future a sulfuric acid plant and fertilizer complex are developed in the area, the values of by-product sulfur and ammonia would be raised.

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* Since this analysis was conducted, the Colorado Legislature has amended the severance tax legislation pertaining to oil shale. While the basic rate for aboveground retorting is unchanged, the various exemptions discussed above are reduced. This will result in plants paying slightly more severance tax, which marginally increases the severance tax credit, thereby marginally (much less than 1%) reducing the pollution control cost.

The by-product value of $32 per barrel for light oils recovered by pollution control activities is higher than the selling price assumed for shale oil, which is $30 per barrel. Light oils are more valuable than heavy oils, and it is the lighter fractions that would be prevented from evaporation by the pollution controls. Consequently, a higher value is justified for recovered shale oil as opposed to whole shale oil.

Working Capital--

The working capital associated with a control was taken as one month's total operating cost plus three months' by-product credit. This is equivalent to be one month's total operating cost disregarding the by-product credit, plus two months' by-product credit. Two months' by-product credit represents one month's inventory and one month's receivables. These values were selected by DRI after review of a variety of data sources.

Working capital is advanced in accordance with the direct annual operating cost plus the extra start-up cost, as follows:

	Operating (On-Stream) Factor	Output as % of Full Production	Operating Cost Relative to Full Production	Working Capital Increment
Year 1	50%	56%	76%	76%
Year 2	75%	83%	83%	7%
Year 3	90%	100%	100%	17%
				100%

Seventy-six percent of the working capital is advanced in Year 1 because this includes the 20% extra start-up cost (56% + 20% = 76%). In Year 2, the operating cost increases from 76% to 83% of normal, hence 7% more working capital is required. A similar argument applies to Year 3, leading to a 17% working capital increment. All working capital is recovered in Year 20.

The working capital charge rate (RW) is calculated in a similar way to a fixed charge rate (see Sections 6.4.1 and 6.4.2). For 12% DCF ROR and normal project-timing assumptions, RW = 20.83%.

6.2.3 Solid Waste Management Costs

Throughout this manual a distinction is made between fixed capital costs and annual operating costs. The importance of this distinction is related to the treatment for determining income tax liability. Operating costs can be claimed as an expense in the year in which they are incurred, whereas a fixed capital cost must be depreciated over the period for which the asset is expected to be used. The effect of classifying a cost as an operating cost rather than a capital cost is to reduce the tax liability in any given year.

For air and water pollution controls, the distinction between fixed capital and annual operating costs is unequivocal. For solid waste management costs which are developed in the form of year-by-year cash flows

(Table 6.1-3), the distinction is less clear. The cost of deep monitoring wells, which occurs only in Year 1 (the first year of production in this case study analysis), was treated as a fixed capital cost, while costs that occur throughout the project life were considered as operating costs. Costs that occur at the end of the project (e.g., revegetation) were also treated as operating costs, since there is no remaining project life over which to depreciate them. In the Lurgi-Open Pit case study, there are two costs, the shallow monitoring wells and piezometers, that occur only in Year 10, i.e., halfway through the project's life. Although by no means a clear-cut decision, these costs were designated fixed capital costs since there is still sufficient time over which to depreciate the assets before the project ends.

Since the solid waste management operating costs are not proportional to production, they were "levelized" to transform them into equivalent direct annual operating costs that are proportional to production, so that they can be treated in the same way as other direct annual operating costs. Levelizing involves determining the annual cost that is proportional to production and which has the same present value (for a given DCF ROR) as the irregular operating cost stream. Further explanation and an example are provided in Section 6.4.3. Costs designated as fixed capital were not levelized.

Table 6.2-3 presents the solid waste management fixed capital costs and direct annual operating costs (levelized at 12% DCF ROR) derived from Table 6.1-3.

6.2.4 Control Cost Example

Table 6.2-4 provides an example of the composition of the various elements of per-barrel control cost for a single major pollution control, the electrostatic precipitators. Per-barrel costs follow identical proportions to annual costs.

It can be seen that the fixed capital charge amounts to 68.4% of the total cost, whereas the working capital charge is only 0.5% of the total cost. It is interesting to note that the fixed capital charge is almost entirely return on equity, as the investment tax credit (20% of fixed capital cost) almost offsets the income tax liability over the project life when both are discounted at 12%, which is the specified DCF ROR. This illustrates the effect of the time-value of money, as the tax credit is given before production commences, whereas the regular tax liability is weighted toward the later years of the project.

The direct operating cost for the electrostatic precipitators is 17.8% of the total cost. Electricity (15.0%) is the largest component, followed by maintenance. This particular pollution control has no operating labor or supplies.

The indirect operating cost amounts to 13.3% of the total cost for this control, of which 12.6% results from the cost of property tax and insurance. The extra start-up costs and the severance tax credit are 2.1% and 1.4%, respectively, of the total.

TABLE 6.2-3. FIXED CAPITAL AND DIRECT ANNUAL OPERATING COSTS FOR SOLID WASTE MANAGEMENT

Activity	Fixed Capital Cost ($000's)	Direct Annual Operating Cost[a] ($000's/yr)
SURFACE HYDROLOGY		
Runoff Collection Sumps		63
Runoff Collection Pumps		16
Deep Monitoring Wells	858[b]	
Shallow Monitoring Wells	26[c]	
Piezometers	432[c]	
SURFACE STABILIZATION		
Dust Suppression		11,079
Revegetation		8
Topsoil		2
Seed		1

[a] The direct annual operating costs are levelized with respect to production at 12% DCF ROR.

[b] Spent in first year of production, Year 1.

[c] Spent in tenth year of production, Year 10.

Source: DRI.

TABLE 6.2-4. PER-BARREL COST BREAKDOWN FOR ELECTROSTATIC PRECIPITATORS
(Standard Economic Assumptions)

Cost Category	Cents/Barrel		Percentage of Total	
Fixed Capital Charge				
Equity Return (12% ROR)	37.4		64.6	
Income Taxes Paid	9.6		16.6	
Investment Tax Credit	(7.4)		(12.8)	
		39.6		68.4
Working Capital Charge		0.3		0.5
Direct Operating Costs				
Maintenance	1.6		2.8	
Operating Supplies	--		--	
Operating Labor	--		--	
Cooling Water	--		--	
Steam	--		--	
Electricity	8.7		15.0	
		10.3		17.8
Indirect Operating Costs				
Taxes and Insurance	7.3		12.6	
Extra Start-up Costs	1.2		2.1	
Severance Tax Credit	(0.8)		(1.4)	
By-product Credit	--		--	
		7.7		13.3
TOTAL COST		57.9		100.0

Source: DRI.

These cost proportions for the electrostatic precipitators are typical of those for air pollution controls. However, for some controls, the indirect operating cost or even the per-barrel control cost can become negative where there is a significant by-product credit.

Water pollution control costs tend to be less capital-intensive, i.e., the ratio of the total annual capital charge to the total annual operating cost is lower. This is because some controls have high operating supplies and utility costs.

Solid waste management costs are different in that they are basically either a fixed capital cost or a direct annual operating cost, but not both for a given control. This reduces working capital and indirect annual operating costs, respectively, to essentially zero.

6.3 COST ANALYSIS RESULTS

The methodology used to develop the data presented in this section is identical to a complete discounted cash flow evaluation; that is, it solves for the annual or per-barrel revenue required to provide the specified return on the investment (DCF ROR) associated with a control. This revenue requirement is known as the total annual or per-barrel control cost. The cost methodology is outlined in Section 6.2, and further details are provided in Section 6.4.1.

Two control items--proper maintenance of valves and pumps and the floating roof oil storage tanks--have relatively large by-product credits which lead to negative total annual costs (i.e., total annual cost credits). Although these items might consequently not be considered pollution controls, their costs have been included in the total cost of air pollution control. The net credit associated with these items represents a very small proportion (less than 0.6%) of the total air pollution control cost using standard economic assumptions, and even less using the sensitivity analyses.

6.3.1 Results for Standard Economic Assumptions*

The term "standard economic assumptions" is used to describe the normal economic assumptions presented in Tables 6.2-1 and 6.2-2. The majority of these assumptions are in reasonable accord with normal engineering and economic evaluation practices. The most critical economic assumption is that

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* As already mentioned, this analysis was developed prior to enactment of the Economic Recovery Tax Act of 1981. The rapid depreciation (ACRS) permitted by this act would significantly reduce the values of the fixed charge factors, especially for normal ("pass through") financing as opposed to stand-alone financing.

For standard economic assumptions, very rough estimates of the changes in total annual control costs are as follows:

Air controls:	10% decrease on aggregate.
Water controls:	5% decrease on aggregate.
Solid waste mgt.:	0-15% decrease, depending on item.

As an alternative assumption, if the energy portion (10%) of the investment tax credit were allowed to expire at the end of 1982, the combined effect of this and ACRS would be to cause small increases in total annual control costs.

of 12% required DCF ROR. This figure was adopted for the oil shale PCTMs and would be appropriate for a mature industry, but it is probably low for a pioneer plant at this time (see Sections 6.2.1 and 6.3.2 for a discussion of factors influencing the selection of a DCF ROR).

Table 6.3-1 provides a detailed summary of pollution control costs, by control group, developed using the standard economic assumptions for the case study considered in this manual. Table 6.3-2 details the specific controls included in each control grouping. Note that total costs for solid waste management are not provided. A complete solid waste management plan for the Lurgi-Open Pit plant has not been proposed. As a result, cost estimates are available for particular items only, and no estimate of total solid waste management cost can be made at this time.

Table 6.3-1 shows that the total fixed capital cost for all air pollution control equipment is approximately $91 million, while the total per-barrel control cost is $1.14. The total fixed capital cost for water pollution control is approximately $7 million, and the total per-barrel control cost is 19 cents.

Table 6.3-1 also compares the per-barrel cost of pollution control to an assumed $30 per-barrel value for shale oil.* For air pollution control, the proportion is 3.8 percent. The total water pollution control cost represents approximately 0.6% of the $30 per-barrel value of shale oil.

The works-gate value of $30 per barrel (mid-1980 dollars) for Lurgi retorted shale oil was based on two sources: a developer's estimate of $29 for a light shale oil (Cathedral Bluffs Shale Oil Co., November 14, 1980), and a study by Peat, Marwick, Mitchell & Co. (September 1980) which derived current values for shale oil. This study concluded that the per-barrel value of shale oil (at the project site) was approximately $31.50 to $32.50 for surface retorted oil. In no case was upgrading involved.

It is generally anticipated that the real price of oil will increase in the future. Hence, the value of $30 may be considered to be a conservative estimate because it does not include any element of future escalation relative to the general level of prices. For example, if oil prices were to escalate at only 2% per annum relative to general cost levels (which can be expected to include pollution control costs), the real value of shale oil would reach almost $45 per barrel (in mid-1980 dollars) by the year 2000, i.e., at the end of the 20-year project life.

Cost Details--

Full cost details for each air and water pollution control (using standard economic assumptions) are presented in Tables 6.3-3 and 6.3-4. As already noted, two items--proper maintenance of valves and pumps and the floating roof oil storage tanks--were found to have negative total annual

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* Other prices for the value of shale oil are used in the other oil shale PCTMs, reflecting quality differences.

TABLE 6.3-1. POLLUTION CONTROL COSTS, BY CONTROL GROUP, FOR THE STANDARD ECONOMIC ASSUMPTIONS

Control Group[a]	Fixed Capital Cost[b] ($000's)	Total Annual Capital Charge[c] ($000's/yr)	Total Annual Operating Cost ($000's/yr)	Total Annual Control Cost ($000's/yr)	Per-barrel Control Cost (cents/bbl)	Per-barrel Control Cost as a Proportion of Oil Value[d] (%)
Air Pollution Control						
Particulate Control	32,451	5,168	4,471	9,639	47	1.5
Flue Gas Treatment	50,734	8,269	3,747	12,016	58	1.9
Miscellaneous Air	7,857	1,310	795	2,105	10	0.3
TOTAL AIR	91,042	14,747	9,013	23,760	115	3.8
Water Pollution Control						
Retort Water	3,788	685	1,745	2,430	12	0.4
Miscellaneous Water	3,334	727	701	1,428	7	0.2
TOTAL WATER	7,122	1,412	2,446	3,858	19	0.6

[a] Refer to Table 6.3-2 for a listing of the items that are included in each control group.

[b] Does not include working capital.

[c] Includes charge for working capital.

[d] Assuming shale oil is valued at $30/barrel.

Source: DRI.

TABLE 6.3-2. CONTROL GROUPINGS

Group Designation	Specific Controls
Air Pollution Control	
Particulate Control:	Fabric filters, water and foam sprays.
Flue Gas Treatment:	Electrostatic precipitators.
Miscellaneous Air:	Stretford, ammonia storage, floating roof oil storage tanks, proper maintenance of valves and pumps, catalytic converters.
Water Pollution Control	
Retort Water:	Ammonia recovery unit, API oil/water separator.
Miscellaneous Water:	Mine water clarifier,* boiler feedwater treatment,* cooling water treatment,* equalization pond, runoff oil/water separator, aeration pond.

* These technologies could be considered as part of the process rather than pollution control.

Source: DRI.

costs. In these cases, the annual by-product credits were large enough to more than offset the total annual capital charges and total annual operating costs. These items were, nevertheless, incorporated into the air pollution control cost total.

Table 6.3-5 presents the costs of nine solid waste management items. Of the nine, dust suppression is by far the most costly item--$11.3 million total annual control cost, or 54 cents per barrel. This item is entirely an operating expenditure (zero fixed capital cost). The only solid waste management items with fixed capital costs are the deep monitoring wells, the shallow monitoring wells, and the piezometers, which total $1.3 million.

It should be remembered that these costs do not represent the full cost associated with a complete solid waste management operation. Even so, the per-barrel control cost associated with these nine solid waste management items is significantly greater than the total per-barrel control cost for water pollution control.

TABLE 6.3-3. DETAILS OF AIR POLLUTION CONTROL COSTS, STANDARD ECONOMIC ASSUMPTIONS

Control Identification (No. of Units)	Fixed Charge Factor (%)	Fixed Capital Cost ($000's)	Working Capital ($000's)	Total Annual Capital Charge[a] ($000's/yr)	Direct Annual Op. Cost ($000's/yr)	Annual By-product Credit ($000's/yr)	Indirect Annual Op. Cost[b] ($000's/yr)	Total Annual Op. Cost ($000's/yr)	Total Annual Control Cost ($000's/yr)	Per-barrel Control Cost (cents)
Fabric Filters (2)	15.61	987	8	156	63	--	31	94	250	1.2
Fabric Filter (1)	15.61	105	1	17	6	--	3	9	26	0.1
Fabric Filter (1)	15.61	552	4	87	34	--	17	51	138	0.7
Fabric Filters (3)	15.61	1,051	8	166	65	--	33	98	264	1.3
Fabric Filters (2)	15.61	628	9	100	85	--	20	105	205	1.0
Fabric Filters (8)	15.61	4,828	38	762	297	--	152	449	1,211	5.8
Fabric Filters (8)	15.61	4,828	38	762	297	--	152	449	1,211	5.8
Fabric Filters (9)	15.61	5,432	42	857	334	--	172	506	1,363	6.6
Fabric Filters (9)	15.61	5,432	42	857	334	--	172	506	1,363	6.6
Fabric Filter (1)	15.61	249	2	39	15	--	8	23	62	0.3
Fabric Filters (2)	15.61	559	4	88	34	--	18	52	140	0.7
Fabric Filters (3)	15.61	559	4	88	34	--	18	52	140	0.7
Fabric Filters (2)	15.61	348	3	55	22	--	11	33	88	0.4
Fabric Filters (2)	15.61	348	3	55	22	--	11	33	88	0.4
Fabric Filters (2)	15.61	348	3	55	22	--	11	33	88	0.4
Fabric Filters (4)	15.61	696	6	110	44	--	22	66	176	0.8
Fabric Filters (2)	15.61	348	3	55	22	--	11	33	88	0.4
Fabric Filters (2)	15.61	348	3	55	22	--	11	33	88	0.4
Water and Foam Sprays	15.61	909	124	168	1,456	--	27	1,483	1,651	8.0
Fabric Filters (4)	16.17	1,837	14	300	113	--	58	171	471	2.3
Fabric Filters (13)	16.17	2,059	16	336	127	--	65	192	528	2.6
Subtotal Particulate Controls		32,451	375	5,168	3,448	--	1,023	4,471	9,639	46.5
Stretford (1)	16.17	6,860	94	1,129	769	72[c]	146	915	2,044	9.9
Ammonia Storage (1)	15.61	466	1	73	--	--	15	15	88	0.4
Floating Roof Storage Tanks (2)	15.61	300	27	52	--	155[d]	(141)	(141)	(89)	(0.4)
Maintenance of Valves, etc.	15.61	61	26	15	61	126[d]	(120)	(59)	(44)	(0.2)
Catalytic Converters	23.36	170	5	41	60	--	5	65	106	0.5
Subtotal Misc. Air Controls		7,857	153	1,310	890	353	(95)	795	2,105	10.2
Electrostatic Precipitators (13)	16.17	50,734	312	8,269	2,144	--	1,603	3,747	12,016	57.9
TOTAL AIR POLLUTION CONTROLS		91,042	840	14,747	6,482	353	2,531	9,013	23,760	114.6

[a] Includes fixed and working capital charges. RW = 20.83%.

[b] Includes annual by-product credit.

[c] For sulfur at $30/long ton.

[d] For light shale oil at $32/bbl.

Source: DRI estimates based on data provided by SWEC.

TABLE 6.3-4. DETAILS OF WATER POLLUTION CONTROL COSTS, STANDARD ECONOMIC ASSUMPTIONS

Control Identification	Fixed Charge Factor (%)	Fixed Capital Cost ($000's)	Working Capital ($000's)	Total Annual Capital Charge[a] ($000's/yr)	Direct Annual Op. Cost ($000's/yr)	Annual By-product Credit ($000's/yr)	Indirect Annual Op. Cost[b] ($000's/yr)	Total Annual Op. Cost ($000's/yr)	Total Annual Control Cost ($000's/yr)	Per-barrel Control Cost (cents)
Ammonia Recovery Unit	16.17	3,627	349	659	2,419	816[c]	(683)	1,736	2,395	11.6
API Oil/Water Separator	16.17	161	1	26	4	--	5	9	35	0.2
Subtotal Retort Water		3,788	350	685	2,423	816	(678)	1,745	2,430	11.8
Mine Water Clarifier[d]	21.64	2,560	33	561	319	--	81	400	961	4.6
Cooling Water Treatment[d]	16.17	--	4	1	51	--	(<1)	51	52	0.3
Boiler Feedwater Treatment[d]	16.17	122	6	21	69	--	4	73	94	0.5
Equalization Pond	21.64	181	1	39	3	--	6	9	48	0.2
Runoff Oil/Water Separator	21.64	41	<1	9	1	--	1	2	11	0.1
Aeration Pond	21.64	430	14	96	153	--	13	166	262	1.3
Subtotal Misc. Water		3,334	58	727	596	--	105	701	1,428	7.0
TOTAL WATER POLLUTION CONTROLS		7,122	408	1,412	3,019	816	(573)	2,446	3,858	18.8

[a] Includes fixed and working capital charges. RW = 20.83%.

[b] Includes annual by-product credit.

[c] For ammonia at $110/ton.

[d] These technologies could be considered as part of the process rather than pollution control.

Source: DRI estimates based on data provided by WPA.

TABLE 6.3-5. DETAILS OF SOLID WASTE MANAGEMENT COSTS, STANDARD ECONOMIC ASSUMPTIONS

Control Identification	Fixed Charge Factor (%)	Fixed Capital Cost ($000's)	Working Capital ($000's)	Total Annual Capital Charge* ($000's/yr)	Direct Annual Op. Cost ($000's/yr)	Indirect Annual Op. Cost ($000's/yr)	Total Annual Op. Cost ($000's/yr)	Total Annual Control Cost ($000's/yr)	Per-barrel Control Cost (cents)
SURFACE HYDROLOGY									
Runoff Collection Sumps	--	--	5	1	63	(<1)	63	64	0.3
Runoff Collection Pumps	--	--	1	<1	16	--	16	16	0.1
Deep Monitoring Wells	12.49	858	2	108	--	27	27	135	0.7
Shallow Monitoring Wells	4.51	26	<1	1	--	1	1	2	<0.1
Piezometers	4.51	432	1	20	--	14	14	34	0.2
SURFACE STABILIZATION									
Dust Suppression	--	--	923	192	11,079	(14)	11,065	11,257	54.3
Revegetation	--	--	1	<1	8	--	8	8	<0.1
Topsoil	--	--	<1	--	2	--	2	2	<0.1
Seed	--	--	--	--	1	--	1	1	<0.1

* Includes fixed and working capital charges. RW = 20.83%.

Note: There are no by-product credits.

Source: DRI estimates based on data provided by SWEC.

6.3.2 Sensitivity Analyses

This section explores the sensitivity of the results to changes in the engineering costs and economic assumptions. In general, only a single change from the standard economic assumptions was made in each case, enabling the impact of this change to be isolated. Table 6.3-6 summarizes the changes made for each case, while Table 6.3-7 displays the fixed and working capital charge rates used to calculate per-barrel control costs. Per-barrel pollution control costs, expressed as a percentage of a $30 per-barrel shale oil value, are given in Table 6.3-8. Table 6.3-9 provides additional detail for the absolute per-barrel control costs and includes percentage changes from the standard economic assumptions. Comparative results for the various sensitivity analyses are presented graphically in Figure 6.3-1. No sensitivity analysis has been performed on the solid waste management costs, as only partial cost estimates were available. Each sensitivity analysis is discussed below.

Twenty Percent Increase in Fixed Capital Costs--

Cost escalation is always a problem with pioneer plants because of the numerous uncertainties (Merrow, September 1978; Merrow, Chapel and Worthing, July 1979). A 20% increase is not at all unreasonable despite the inclusion of a 20% contingency in fixed capital cost estimates.

Table 6.3-9 shows that a 20% increase in fixed capital costs has a moderate effect on pollution control costs. As would be expected, the more capital-intensive air pollution controls show the greatest increase. The total air pollution control cost increases by 15% (16 cents per barrel), while the total water pollution control cost increases by 8% (only 1 cent per barrel).

Twenty Percent Increase in Operating Costs--

Operating costs are often better defined than capital costs, which is why an operating cost contingency is not normally included in the direct annual operating costs. However, there are many reasons why operating costs could be higher than anticipated. For example, regional shortages of skilled labor could result in higher wages and reduced productivity. Also, labor costs may escalate faster than other costs. Maintenance costs could be higher than expected, and both utility requirements and utility unit costs could deviate from expectations.

For air pollution controls, the overall effect of an increase in direct annual operating cost is much less than that of the same percentage increase in fixed capital cost. For a 20% increase, the total air pollution control cost increases by only 6 cents per barrel (a 6% increase). The more operating cost-intensive total water pollution control cost increases by 3 cents per barrel (a 16% increase). This is a reversal of the results obtained for a 20% increase in fixed capital costs, and confirms that the air pollution controls are much more capital-intensive than the water pollution controls.

TABLE 6.3-6. ASSUMPTIONS FOR SENSITIVITY ANALYSES*

Sensitivity Analysis	DCF ROR	Fixed Capital Costs	Direct Operating Costs	By-product Credits	Comments
+20% Fixed Capital Costs	12%	Increased 20%	SEA	SEA	
+20% Direct Operating Costs	12%	SEA	Increased 20%	SEA	
+66.7% Utilities Costs	12%	SEA	Utility portion increased 66.7%	SEA	
80% of Planned Output	12%	SEA	Decreased 10%	Decreased 20%	
Delayed Start-up	12%	SEA	SEA	SEA	A 2-year delay was incorporated into the RC and RW calculations by halting production in Years 2 and 3 and resuming in Year 4. Project life was increased to 22 years.
15% DCF ROR	15%	SEA	SEA	SEA	
Stand-alone Financing	12%	SEA	SEA	SEA	For RC and RW calculations, investment tax credit and depreciation earned in or before Year 3 were accumulated and taken as a lump sum in Year 3. The schedules after Year 3 remained unchanged.
Stand-alone Financing at 15% DCF ROR	15%	SEA	SEA	SEA	
+20% Fixed Capital Costs, Delayed Start-up and 15% DCF ROR	15%	Increased 20%	SEA	SEA	A 2-year delay was incorporated into the RC and RW calculations by halting production in Years 2 and 3 and resuming in Year 4. Project life was increased to 22 years.
+20% Fixed Capital Costs, Delayed Start-up, 15% DCF ROR and Stand-alone Financing	15%	Increased 20%	SEA	SEA	A 2-year delay was incorporated into the RC and RW calculations as above, and the investment tax credit and depreciation were accumulated to Year 5.

* SEA indicates that the costs are the same as those used for analysis based on standard economic assumptions.

Source: DRI.

TABLE 6.3-7. CHARGE RATES FOR SENSITIVITY ANALYSES

	Standard Economic Assumptions	Sensitivity Analyses									
		+20% Fixed Capital Costs	+20% Direct Operating Costs	+66.7% Utilities Costs	80% of Planned Output	Delayed Start-up	15% DCF ROR	Stand-alone Financing	Stand-alone Financing at 15% DCF ROR	Combined Assumptions[a]	Combined Assumptions with Stand-alone Financing[a]
Fixed Charge Rate[b]											
Retort Timing	16.17	16.17	16.17	16.17	16.17	19.92	20.92	18.52	24.31	26.94	33.92
Mine Timing	15.61	15.61	15.61	15.61	15.61	19.23	20.06	17.81	23.23	25.82	32.53
Early Water Management	21.64	21.64	21.64	21.64	21.64	26.66	30.01	26.82	37.62	38.63	51.89
Catalytic Converters	23.36	23.36	23.36	23.36	23.36	26.46	27.03	25.14	29.53	31.91	38.91
Working Capital Charge Rate	20.83	20.83	20.83	20.83	20.83	20.96	25.58	20.83	25.58	25.80	25.80

[a] Combined assumptions are 20% increase in fixed capital costs, 15% DCF ROR and delayed start-up.

[b] Refer to Table 6.2-2 for pollution controls included in each category.

Source: DRI.

TABLE 6.3-8. SENSITIVITY ANALYSES EXPRESSED AS A PERCENTAGE OF SHALE OIL VALUE

Sensitivity Analysis	Per-barrel Control Cost as a Percent of $30/Barrel Shale Oil Value: Air	Water
Standard Economic Assumptions	3.8	0.6
20% Increase in Fixed Capital Costs	4.4	0.7
20% Increase in Direct Operating Costs	4.0	0.7
66.7% Increase in Utilities Costs	4.2	0.8
80% of Planned Output	4.7	0.7
Delayed Start-up	4.4	0.7
15% DCF ROR	4.5	0.7
Stand-alone Financing	4.2	0.7
Stand-alone Financing at 15% DCF ROR	5.0	0.8
Combined Assumptions*	6.3	0.9
Combined Assumptions with Stand-alone Financing*	7.5	1.0

* Combined assumptions are 20% increase in fixed capital costs, 15% DCF ROR and delayed start-up.

Source: DRI.

66.7% Increase in Utilities Costs--

Operation of various controls requires inputs of electricity and steam. Under standard economic assumptions, electricity is valued at 3 cents per kW-hr, and it is assumed that steam is generated at a cost of $3/MMBtu. The electricity charge of 3 cents per kW-hr may very likely underestimate the true cost of power purchased from the grid (should this prove necessary) as it is a compromise value between plants that can sell power and those that must purchase power (see Section 6.1.1). Since the Lurgi-Open Pit plant is likely to require electricity from outside sources, a 5 cents per kW-hr rate (a 66.7% increase) was considered. At the same time, the cost of steam was also increased by 66.7%, as the standard rate for this input of $3/MMBtu may also prove to be conservative. Three dollars per million Btu is a typical 1980 value used for heat inputs in engineering studies, but no detailed cost evaluation was conducted for this manual. Hence, the steam cost must be considered uncertain.

TABLE 6.3-9. SENSITIVITY ANALYSES BY MEDIUM

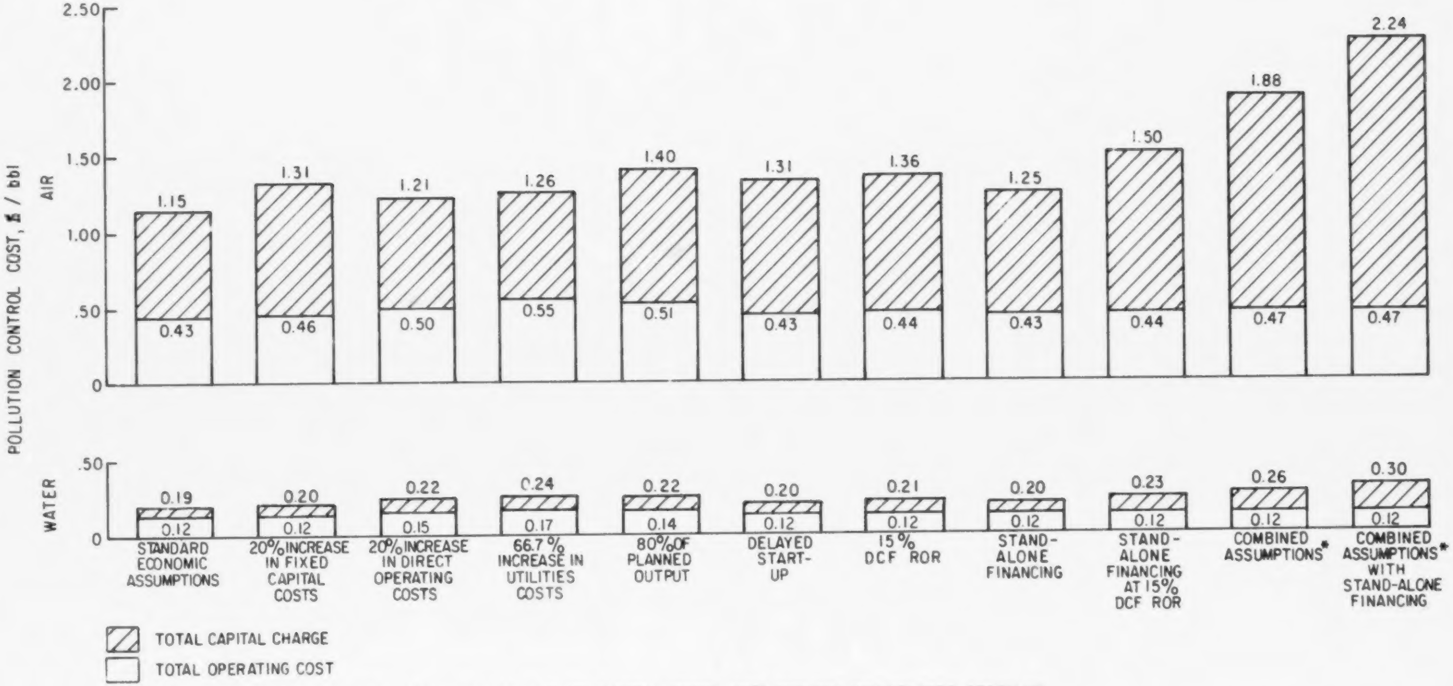
Sensitivity Analysis	Air Pollution Control		Water Pollution Control	
	cents/bbl	% change	cents/bbl	% change
Standard Economic Assumptions	115	--	19	--
20% Increase in Fixed Capital Costs	131	+14.8	20	+8.1
20% Increase in Direct Operating Costs	121	+5.6	22	+16.0
66.7% Increase in Utilities Costs	126	+10.2	24	+28.6
80% of Planned Output	140	+22.0	22	+20.2
Delayed Start-up	131	+14.3	20	+8.1
15% DCF ROR	136	+18.5	21	+12.9
Stand-alone Financing	125	+8.8	20	+6.7
Stand-alone Financing at 15% DCF ROR	150	+31.2	23	+22.7
Combined Assumptions*	188	+64.2	26	+39.4
Combined Assumptions with Stand-alone Financing*	224	+95.9	30	+61.1

* Combined assumptions are 20% increase in fixed capital costs, 15% DCF ROR and delayed start-up.

Note: Percentage changes may not agree with figures calculated from cents per barrel due to rounding.

Source: DRI.

The results indicate that utility costs constitute a moderately important component of pollution control costs. The total water pollution control cost increases by 28% (5 cents per barrel). This increase can be attributed to the large quantities of steam required by the ammonia recovery unit. The effect on air pollution control costs is less significant (although the absolute increase in costs is greater). The 66.7% increase in utilities costs causes the total air pollution control cost to rise by 10% (11 cents



* COMBINED ASSUMPTIONS ARE 20% INCREASE IN FIXED CAPITAL COSTS, 15% DCF ROR AND DELAYED START-UP.

SOURCE: DRI

FIGURE 6.3-1 SENSITIVITY ANALYSES: TOTAL PER-BARREL AIR AND WATER POLLUTION CONTROL COSTS

per barrel). This increase is due largely to the significant amounts of electricity required to operate the electrostatic precipitators and fabric filters.

Eighty Percent of Planned Output--

A frequent problem with pioneer process plants is that they fail to achieve their planned output. Occasionally they produce more. When a plant fails to reach its planned output, the annual fixed capital charges must be spread over reduced output, and the direct annual operating costs decrease by a lesser proportion than the output because some components (such as maintenance) are virtually unchanged.

For the case of a plant that achieves only 80% of planned output, it was assumed that direct annual operating costs fall to 90% of the full production costs. Production in the start-up years and by-product credits were prorated to 80% of the standard values.

Overall, the results are relatively severe, with the more capital-intensive air pollution controls showing the greatest increase. Total air pollution control cost increases 22% (25 cents per barrel), while the total water pollution control cost increases 20% (3 cents per barrel).

Delayed Start-up--

Because of the time-value of money implicit in the discounting procedure, anything that delays or curtails production raises annual capital charges and, hence, the per-barrel control cost; conversly, anything that accelerates or extends production reduces the costs.

For this analysis, production is halted for two years (Years 2 and 3) and then follows the normal build-up profile displaced by two years. (The project life is extended by 2 years to 22 years.) This profile corresponds to the scenario that the plant initially starts production according to schedule; then, at the end of Year 1, the plant is closed down because serious operational problems have developed and must be solved, which takes two years.

The effects of this case are only moderately severe. Total air pollution control cost increases 14% (16 cents per barrel). The less capital-intensive total water pollution control cost increases by 8% (1 cent per barrel).

Fifteen Percent DCF ROR--

The minimum acceptable DCF ROR used in a project feasibility study is normally not divulged by developers and, in any event, is influenced by alternative investment opportunities and other factors. However, there is broad confirmation that a rate between 12% and 15% per annum (in constant dollars) is appropriate for evaluating oil shale investments (Denver Research Institute, et al., July 1979; also see Merrow, September 1978). This ROR, which is called a "hurdle rate," is higher than the return that a company actually earns on its capital for a number of reasons. First, it is an unfortunate fact of life that many projects earn less than the projected rate because things do not work out as expected. This is only partly offset by the few that do better than anticipated. Second, project evaluations do not usually include such costs as R and D, exploration, and reserve acquisition; also, they may not include recovery of some general corporate expenses.

The single most important factor that influences the required DCF ROR is the perceived riskiness of the project. A high risk project is expected to pass a higher ROR hurdle than a low risk project. Some of the types of risks that might be subjectively taken into account in selecting a minimum acceptable ROR for a mining project in the U.S. include:

- Unproven technology (and, hence, uncertain equipment costs);
- Geologic uncertainty;
- Very large investments in relationship to total corporate assets;
- Rapid inflation in some cost components;
- Long construction and start-up periods;
- Market uncertainty;
- Regulatory uncertainty (leading to delays or added costs); and
- Difficult working conditions or adverse socioeconomic impacts leading to manpower problems.

For any first generation commercial synfuel plant, all the above factors are present, with the possible exception of geologic uncertainty. At this time, most of these factors are strongly present in oil shale projects. The standard economic assumption is 12% DCF ROR, which is probably the lowest acceptable ROR for a private enterprise shale oil plant with proven technology. For a pioneer plant, industry is likely to require at least 15% ROR, unless it wishes to "buy into" a new industry. Of course, if another party (e.g., the Federal government) were prepared to share the risk in some way, the required ROR would be reduced. Even though some of the risks listed above do not apply to pollution controls, industry does not perceive environmental costs to be separable from the total project. Hence, all components of a project, including pollution controls, must earn the specified DCF ROR.

Increasing the required DCF ROR from 12 to 15% has a substantial effect on pollution control costs. Once again, air pollution controls show the greatest increase. The total air pollution control cost increases by 19% (21 cents per barrel), while the total water pollution control cost increases by 13% (or 2 cents per barrel).

Stand-alone Financing--

The term "stand-alone financing" is used to describe a project in which investment tax credits and allowances for depreciation cannot be passed through to a parent company (or companies) which can benefit from them immediately. (These benefits are treated as negative income tax in conducting the alternative "pass-through" form of project evaluation which is used under standard economic assumptions.) Instead, it is necessary for the project to become profitable before the tax benefits can be obtained. It is difficult to determine when this might occur because it requires a detailed knowledge of the overall project economics; in any event, the timing of the benefits will be affected by the selling price of the shale oil. However, it is known that some of the developers are assuming stand-alone financing for their evaluations since it more closely reflects their tax positions than does pass-through financing.

To determine the approximate effect of substituting stand-alone financing for pass-through financing, it was assumed that no investment tax credit or depreciation could be claimed until the third year of production, i.e., the first year of full output. This assumption was based on examination of the cash flow analysis for an open pit mine with surface retorting presented in a recent oil shale tax study (Peat, Marwick, Mitchell & Co., September 1980). It must be emphasized that this assumption is very simplistic (and probably conservative), since the relevant details in the tax study were significantly different from those assumed in this manual. As expected, the effect was larger for the more capital-intensive air pollution controls, although the overall effect for both control groups is fairly mild. Total air pollution control cost increases 10 cents per barrel (9%), while the total water pollution control cost increases 1 cent per barrel (7%). A more refined calculation might yield substantially greater increases, especially if a low value was used for the price of shale oil, thereby reducing profitability.

The effect of stand-alone financing was also evaluated at 15% DCF ROR, using the same assumptions as above. This probably comes closer to a developer's evaluation. The resulting increases in costs are quite substantial, with the total air pollution control cost increasing 35 cents per barrel (31%) and the total water pollution control cost increasing 4 cents per barrel (23%).

Combined Cases--

Two combined cases were evaluated using the components already discussed. However, it is not sufficient to construct these analyses by simply combining the results from the earlier findings, so new analyses were developed. The two cases are as follows:

Combined assumptions

- 20% increase in fixed capital costs
- Delayed start-up
- 15% DCF ROR
- Everything else as standard economic assumptions.

Combined assumptions with stand-alone financing

- 20% increase in fixed capital costs
- Delayed start-up
- 15% DCF ROR
- Stand-alone financing
- Everything else as standard economic assumptions.

These combined cases are intended to be quite plausible adverse scenarios (i.e., 20% increase in fixed capital costs and delayed start-up) looked at from industry's viewpoint (i.e, 15% DCF ROR, with or without stand-alone financing, depending on the company).

The results indicate that these cases would impose significant burdens on industry. The more capital-intensive air pollution controls increase in cost by 64% (73 cents per barrel) for regular ("pass-through") financing and by 96% ($1.09 per barrel) for stand-alone financing. Total water pollution control cost rises approximately 39% (7 cents per barrel) for the regular case and 61% (11 cents per barrel) for the stand-alone case. The absolute level of pollution control costs reaches $1.88 per barrel for all air controls and 26 cents per barrel for water pollution controls for the regular (pass-through) case. For combined assumptions with stand-alone financing, absolute pollution control costs are $2.24 per barrel for total air and 30 cents per barrel for total water. These results represent an almost doubling of the absolute cost of air pollution controls.

Summary--

Returning to Table 6.3-8, it can be seen that the total cost of air pollution control is roughly 4% of the assumed $30 per-barrel value for shale oil under the standard economic assumptions. The total water pollution control cost is roughly 0.6% of the value of the oil.

With respect to air pollution controls, only the two sets of combined assumptions produce major increases in cost. In these two cases, the total control cost reaches 6.3 and 7.5% of the assumed $30 value for shale oil.

Water pollution control costs have proven to be less sensitive to changes in the engineering costs and economic assumptions. Only the last two sensitivity analyses (the two sets of combined assumptions) produce noticeable increases in total water pollution control costs. From a base of 0.6% of the shale oil value under the standard economic assumptions, water pollution control cost rises no higher than to 1.0% of the oil value (for combined assumptions with stand-alone financing). When compared with air pollution control costs, water control costs are more sensitive to changes in direct operating costs and utilities, as opposed to changes that affect fixed capital charges. Increases in direct operating costs and utilities costs, however, do not produce significantly larger increases in total water pollution control costs than those sensitivity analyses which affect fixed capital charges.

Figure 6.3-1 splits the pollution control costs into a per-barrel total capital charge and a per-barrel total operating cost. This figure effectively illustrates the response of capital-intensive controls (air) vs. operating cost-intensive controls (water) to the different sensitivity analyses.

6.4 DETAILS OF COST ANALYSIS METHODOLOGY

6.4.1 Cost Algorithms

This section provides the algorithms used to calculate total annual and per-barrel control costs and capital charge factors.

Calculation of Total Annual and Per-barrel Control Costs--

The total annual control cost (TC) of each item considered for pollution control is the sum of the total annual operating cost (TOC) and the total annual capital charge (CC). That is:

$$TC = TOC + CC$$

and $$TOC = DOC + IOC$$

where:

DOC = Direct annual operating cost
IOC = Indirect annual operating cost

and $$CC = (FCC \times RF) + (WC \times RW)$$

where:

FCC = Fixed capital cost
WC = Working capital
RF = Fixed charge factor
RW = Working capital charge factor

The cost per barrel (CPB) is the total annual cost divided by the normal annual production, i.e.:

$$CPB = TC \div (BPSD \times 328.5)$$

where: BPSD = Barrels per stream day

The factor, 328.5, is the number of normal operating days per year.

The derivation of each cost component is explained below.

Direct annual operating cost. DOC is a data input derived from the engineering cost analysis. It is the annual cost for a normal year and is taken from one of the data Tables 6.1-1, 6.1-2 or 6.2-3.

Indirect annual operating cost. The indirect annual operating cost (IOC) is calculated as follows:

$$IOC = TIA + ESC - STC - BP$$

where:
TIA = Annual property tax and insurance allowance
ESC = Annual extra start-up costs (levelized--see below)
STC = Annual severance tax credit (levelized--see below)
BP = Annual by-product credit

BP is an input generated from stream data and shown in one of the tables in Section 6.3, and:

$$TIA = 0.03 \times FCC$$

$$ESC = (0.03 \times FCC + 0.20 \times DOC) \times LFAC1$$

$$STC = 0.04 \times [(DOC + ESC + TIA - BP) + 0.05 \times FCC] \times LFAC2$$

LFAC1 and LFAC2 are levelizing factors that spread ESC and STC uniformly over all units of production. LFAC2 also makes adjustments for the severance tax exemptions allowed for low production. These factors are as follows:

$$LFAC1 = \frac{(1 + r)^{-1}}{\dfrac{0.56}{1 + r} + \dfrac{0.83}{(1 + r)^2} + \sum_{n=3}^{20} \dfrac{1}{(1 + r)^n}}$$

$$= \frac{(1 + r)^{-1}}{\dfrac{0.56}{1 + r} + \dfrac{0.83}{(1 + r)^2} + \dfrac{(1 + r)^{-2} - (1 + r)^{-20}}{r}}$$

$$LFAC2 = \frac{BPSD - 10,000}{BPSD} \times$$

$$\frac{\dfrac{1}{4} \times \dfrac{0.83}{(1 + r)^2} + \dfrac{1}{2} \times \dfrac{1}{(1 + r)^3} + \dfrac{3}{4} \times \dfrac{1}{(1 + r)^4} + \dfrac{(1 + r)^{-4} - (1 + r)^{-20}}{r}}{\dfrac{0.56}{1 + r} + \dfrac{0.83}{(1 + r)^2} + \dfrac{(1 + r)^{-2} - (1 + r)^{-20}}{r}}$$

where:
r = Discount rate = DCF ROR
BPSD = Barrels per stream day (i.e., normal daily output)

A numerical example of a levelizing calculation is given in Section 6.4.3.

Capital costs. Fixed capital cost (FCC) is an input taken from one of the data tables. Working capital (WC) is calculated as follows:

$$WC = 1/12 \times TOC + 1/4 \times BP$$

Capital Charge Factors--

The fixed charge factor equation is:

$$RF = \frac{\sum_{n=J}^{N} [(1 + r)^{-n} \times (K_n - T \times D_n - C_n)]}{(1 - T) \sum_{n=1}^{N} [(1 + r)^{-n} O_n]}$$

where: K_n = Capital expenditure in year n ($\Sigma K_n = 1.000$)
C_n = Investment credit in year n
D_n = Depreciation in year n
O_n = Operating income in year n ($O_n = 1.000$ in a normal year)
r = Discount Rate = DCF ROR
T = Tax rate
N = Last year of project
J = First year of project (i.e., -3)

Note that the first year of production is Year 1.

The same equation is used to determine the working capital charge factor (RW), except that the D_n and C_n terms are omitted.

6.4.2 Example Calculation of a Fixed Charge Factor

Table 6.4-1 provides an example of the calculation of a fixed charge factor. The data used are for retort timing, using standard economic assumptions (see Table 6.2-2).

The following is an explanation of the calculations in the table. Expenditures are shown negative, while income (and taxes avoided) is shown positive. Column [2] is a schedule of capital expenditures to be made over a three-year period, totaling an arbitrary $1,000. (Unit value is used instead of $1,000 in the equation above.) Columns [3], [4], and [5] deal with allowances associated with this capital expenditure. Column [3] is a schedule of depreciation, commencing in Year 1 when the asset is placed into service. Column [4] gives the value of the depreciation allowed to the company. This value is the income tax not incurred as a consequence of the depreciation deduction, and it is 48% of Column [3]. Column [5] is the 20% investment tax credit available in each year a capital expenditure is made. (This is a direct credit against tax and does not have to be multiplied by the tax rate.)

Column [6] represents the income stream resulting from the $1,000 investment (Column [2]). Income in a normal, full production year is

TABLE 6.4-1. EXAMPLE OF FIXED CHARGE FACTOR CALCULATION

(Standard Economic Assumptions, Retort Timing)

		Allowances			Operating Income			Net Present Values			
Year [1]	Gross Capital [2]	Depreciation Amount [3]	Depreciation Value @ 48% Tax [4]	Investment Tax Credit [5]	Gross [6]	Net After 48% Tax [7]	Discount Factors at 12% [8]	After-tax Income* [9]	Depreciation Allowance [10]	Investment Tax Credit [11]	Capital [12]
-2	(100.00)	0.00	0.00	20.00	0.00x	0.00x	1.2544	0.0000x	0.00	25.09	(125.44)
-1	(300.00)	0.00	0.00	60.00	0.00x	0.00x	1.1200	0.0000x	0.00	67.20	(336.00)
0	(600.00)	0.00	0.00	120.00	0.00x	0.00x	1.0000	0.0000x	0.00	120.00	(600.00)
1	0.00	117.65	56.47	0.00	0.56x	0.291x	0.8929	0.2600x	50.42	0.00	0.00
2	0.00	110.29	52.94	0.00	0.83x	0.432x	0.7972	0.3441x	42.20	0.00	0.00
3	0.00	102.94	49.41	0.00	1.00x	0.520x	0.7118	0.3701x	35.17	0.00	0.00
4	0.00	95.59	45.88	0.00	1.00x	0.520x	0.6355	0.3305x	29.16	0.00	0.00
5	0.00	88.24	42.36	0.00	1.00x	0.520x	0.5674	0.2951x	24.03	0.00	0.00
6	0.00	80.88	38.82	0.00	1.00x	0.520x	0.5066	0.2634x	19.67	0.00	0.00
7	0.00	73.53	35.29	0.00	1.00x	0.520x	0.4523	0.2352x	15.97	0.00	0.00
8	0.00	66.18	31.77	0.00	1.00x	0.520x	0.4039	0.2100x	12.83	0.00	0.00
9	0.00	58.82	28.23	0.00	1.00x	0.520x	0.3606	0.1875x	10.18	0.00	0.00
10	0.00	51.47	24.71	0.00	1.00x	0.520x	0.3220	0.1674x	7.95	0.00	0.00
11	0.00	44.12	21.18	0.00	1.00x	0.520x	0.2875	0.1495x	6.09	0.00	0.00
12	0.00	36.76	17.64	0.00	1.00x	0.520x	0.2567	0.1335x	4.53	0.00	0.00
13	0.00	29.41	14.12	0.00	1.00x	0.520x	0.2292	0.1192x	3.24	0.00	0.00
14	0.00	22.06	10.59	0.00	1.00x	0.520x	0.2046	0.1064x	2.17	0.00	0.00
15	0.00	14.71	7.06	0.00	1.00x	0.520x	0.1827	0.0950x	1.29	0.00	0.00
16	0.00	7.35	3.53	0.00	1.00x	0.520x	0.1631	0.0848x	0.58	0.00	0.00
17	0.00	0.00	0.00	0.00	1.00x	0.520x	0.1456	0.0757x	0.00	0.00	0.00
18	0.00	0.00	0.00	0.00	1.00x	0.520x	0.1300	0.0676x	0.00	0.00	0.00
19	0.00	0.00	0.00	0.00	1.00x	0.520x	0.1161	0.0604x	0.00	0.00	0.00
20	0.00	0.00	0.00	0.00	1.00x	0.520x	0.1037	0.0539x	0.00	0.00	0.00
	(1,000.00)	1,000.00	480.00	200.00				3.6093x	265.48	212.29	(1,061.44)

* After-tax income is before depreciation allowance and investment tax credit.

Source: DRI.

designated by "1.00x." Since income is proportional to production, and production in the start-up years is less than full production, the first two years of income are appropriately reduced, i.e., 0.56x in Year 1 (0.56 is the 50% operating factor in Year 1 divided by the 90% factor for a normal year) and 0.83x in Year 2. Column [7] shows the residual income to the company after income tax is paid on the income in Column [6].

The 12% discount factors in Column [8] are used to generate the present values in Columns [9], [10], [11] and [12]. After summing the columns of present values of after-tax income, depreciation allowance, investment tax credit, and capital expenditure, an equation is constructed to determine the gross income, x, which must be generated by the $1,000 of invested capital to achieve a 12% DCF ROR; thus:

$$3.6093x = 1{,}061.44 - 265.48 - 212.29$$

$$[9] = [12] - [10] - [11]$$

therefore: $$x = \frac{583.67}{3.6093} = 161.71$$

(x represents the gross income in a full production year that is necessary to provide the specified DCF ROR, 12%, on $1,000 of fixed capital.)

hence: $$RF = \frac{161.71}{1{,}000} = 16.17\%$$

6.4.3 Cost Levelizing Calculations

While most direct operating costs vary in proportion to plant output, the operating costs for solid waste management do not. A prime example of this is the cost of surface reclamation, which only occurs at the end of the project. To spread these costs in a pattern consistent with production, these operating costs are transformed into an annual figure which can then be applied to each barrel of shale oil produced. This is done by calculating a "levelized cost" for a normal year's production. This technique is also used to spread the extra start-up cost and severance tax credit uniformly over shale oil production.

A "levelizing factor" is used to make this transformation. The following equation shows how a levelizing factor is used to arrive at a levelized cost (i.e., a stream of payments having the same profile as production), given the present value of a nonuniform stream of payments:

$$\text{Levelized Cost} = \frac{\Sigma(\text{Present Values of a Cost Stream})}{\text{Levelizing Factor}}$$

By dividing the levelized cost by a normal year's output, a cost per unit of production is derived.

The equation for calculating the levelizing factor (LF) is:

$$LF = PVFA_{(r,N)} - \sum_{n=1}^{S} (PVF_{(r,n)} \times [1-L_n])$$

where: LF = Levelizing factor

$PVFA_{(r,N)}$ = Present value factor of a uniform series of payments for N years

$PVF_{(r,n)}$ = Present value factor of a single payment in year n

r = Discount Rate = DCF ROR

N = Number of production years

S = Number of years in the start-up period

n = Any specific year in the start-up period

L_n = The proportion of normal output during any given start-up year; the series of L_n values constitutes the "start-up profile"

The second term on the right-hand side of the above equation is an adjustment to the uniform series represented by the first term. The complement of the L_n figure (i.e., that portion of each start-up year which is less than full production) is discounted, summed, and then subtracted from the uniform series. Since the start-up years have high present values, the effect of subtracting this term has a substantial impact on the levelizing factor. Because the levelizing factor is the denominator in the equation which determines the levelized cost (and, hence, the unit cost), this adjustment term raises the per-barrel cost.

__Cost Levelizing Example__--

To illustrate the concept of cost levelization, calculation of the 12% DCF ROR levelizing factor used in this manual is presented below:

Year	Proportion of Normal Output (L_n)	PVF @ 12%	$(1-L_n) \times PVF$
1	0.56	0.8929	0.3929
2	0.83	0.7972	0.1355
3	1.00	5.7793	0.0000
.	.		
.	.		
.	.		
20	1.00		
		7.4694	0.5284

Hence: $LF_{(r=12\%,\ N=20\ yrs)} = 7.4694 - 0.5284 = 6.9410$

(Note that all present values are expressed with respect to Year 0)

This factor is the same as the denominator in the levelizing expressions LFAC1 and LFAC2.

As an illustration of a levelizing calculation, consider the revegetation costs shown in Table 6.1-3. These costs are incurred as follows:

Year 19:	$185,000
Year 20:	$185,000
Year 21:	$185,000

The present value of these costs, expressed with respect to Year 0, is calculated as follows:

Year	Expenditure	PVF @ 12%	Present Values
19	$185,000	0.1161	$21,478
20	185,000	0.1037	19,185
21	185,000	0.0926	17,131
			$57,794

Thus, $57,794 is the present value of all the revegetation costs. To turn this into a cost that is distributed uniformly with respect to output, it must be divided by $LF_{(r=12\%,\ N=20\ \text{years})}$.

$$\text{Therefore, Levelized Cost} = \frac{57,794}{6.9410} = \$8,326$$

Thus, $8,326 (rounded to $8,000 in Table 6.2-3) is the annual cost, in a normal production year, that is equivalent to the irregular cost profile given above. This direct annual operating cost can be used in conjunction with the algorithms given in Section 6.4.1 for calculation of total annual control cost and per-barrel control cost, whereas the irregular stream of expenditures from which it was derived could not be used with the standard methodology.

In summary, cost levelization redistributes a cost series that is not proportional to production in such a way as to yield an equivalent series that is proportional to production and has the same economic value.

SECTION 7

DATA LIMITATIONS AND RESEARCH NEEDS

A number of limitations associated with stream characterization and pollution control technology performance were identified in the data base during the preparation of the Pollution Control Technical Manual for the Lurgi oil shale retorting process combined with open pit mining. It is important that users of this manual be aware of these limitations. It is also important that these limitations be addressed prior to development of an oil shale facility of the magnitude analyzed in this manual (e.g., 119,000 TPSD oil shale mined, 193,000 TPSD total solids mined, and 63,140 BPSD shale oil produced).

7.1 DATA LIMITATIONS

The description of the Lurgi retorting process and information regarding applicable control technologies, performance, and costs used to prepare this manual were obtained from reports on the operation of pilot Lurgi retorts, vendor descriptions, and engineering calculations used in conjunction with experience transferred from analogue industries such as the petroleum, utility, and mineral mining industries which utilize similar control technologies. Until "hands on" experience is obtained from commercial-scale oil shale operations, these sources constitute the best available data base. However, the limitations of this data base should be clearly understood. Pilot retorts were built and operated primarily to improve process design and not for demonstrating operation of a commercial-sized retort with attendant pollution control systems. Many pollution control systems have never been pilot tested with an oil shale retort. Even for those control systems that were pilot tested, often the data collected have been very limited.

The primary experience with Lurgi retorting involves two pilot plants (5 tons/day and 25 tons/day) and several laboratory-scale retorts operated in West Germany during the past few years. Shales from Tract C-a, Tract C-b, and the Colony mine in Colorado have been processed recently, and the available data from these tests have been used in this manual. A full-sized Lurgi retort is expected to process 8,800 TPSD of raw shale, and 13 of these retorts will be needed to produce 63,140 BPSD of shale oil. This represents an enormous scale-up of the pilot retorts; therefore, improvements in the retort design and operating parameters may be inevitable, resulting in some uncertainty about the stream compositions and performance of control technologies.

Variations in the grade of the shale also introduce modifications to the operating parameters and, hence, the data. This is evident from the retorting tests on the oil shale from Tracts C-a and C-b, from which significantly different results were obtained. Thus, a linear extrapolation of the data from these operations may not be entirely applicable to the processing of shales from other locations, and a direct transfer of the information to other development sites must be made with caution.

It should also be noted that, to date, the Lurgi pilot plants have consisted of the retort and flue gas discharge system only. Other unit processes (e.g., oil and gas recovery, naphtha recovery, retort gas compression) and control technologies (e.g., Stretford, ammonia recovery) that form the basis for the complete plant analyzed in this manual have not yet been tested with the Lurgi process. Therefore, actual control technology performance and compatibility with the Lurgi retorting process have not been demonstrated.

The fact that the processing streams have been measured in terms of major constituents only is an additional limitation. Information on minor constituents, which may be of concern from an operational as well as an environmental viewpoint, is not well documented. Examples of such constituents include regulated and nonregulated pollutants (e.g., trace elements, specific organics, inorganics), all of which can have an impact upon the choice and operation of downstream control.

The open pit mining and backfilling operation are also prominent features of this study. Mining a total of 193,000 TPSD of the material will constitute the largest mining operation in the world. The issues associated with the magnitude of this effort are further complicated due to the interception of two aquifers. More mine water may be produced than needed by the plant, which may necessitate the disposal of the excess water. Environmental implications of surface discharge or underground injection of the excess water are not fully known. Backfilling the pit with the plant wastes and reestablishing the aquifers at the end of the project also deserve a thorough analysis at the onset of the project.

Assessing the limitations of existing data sources was an important by-product resulting from the preparation of this manual. Since the best available information on each subject was selected, this manual represents the best currently available data base on the Lurgi retorting and open pit mining processes; also, within the limitations of available data, it accurately estimates the control efficiencies achievable.

7.2 RESEARCH NEEDS

The limited potential for the transfer of control technology from pilot and semi-works retorting tests and from analogue industries to commercial oil shale operations emphasizes a genuine need for research in certain areas of oil shale processing and pollution control. This need is strengthened by the fact that, even with several years of experience, the oil shale industry is still in an early state of development.

While it is recognized that further research will be essential in all phases of oil shale commercialization, the major areas of data uncertainty regarding characterization of streams and control technology performance, as revealed during preparation of the Lurgi-Open Pit PCTM, are identified in Table 7.1-1. The status of the information is presented according to the development stage of the source and technology. The specific information sources are also identified. A reliability or confidence ranking is assigned to the data for each stream and technology based on a subjective evaluation of the direct applicability of the data to a commercial-scale Lurgi-Open Pit facility. Some salient features and caveats in the information base are noted, and specific research needs are identified to overcome some of the data limitations.

TABLE 7.1-1. DATA LIMITATIONS AND RESEARCH NEEDS

Streams and Control Technologies (Figure No.)	Pollutant Controlled	Information Status[a]	Information Sources[b]	Reliability[c]	Remarks	Research Needs
Particulate Emissions (3.3-2, 3.3-10)	--	I	10	2	The particulate emission estimates have been calculated by using dust emission factors for different materials handling operations. Actual source testing is not documented.	The particulate emission data from actual source testing need to be obtained.
					Only the bulk particulates have been estimated. Trace elements and other criteria pollutants have not been estimated.	Data on trace elements and other criteria pollutants need to be obtained from actual source testing.
Baghouses (3.3-2)	Particulates (point source)	I	1	2	The technology is mentioned in the original Tract C-a DDP as the control for point source particulates, but the operating experience with the Lurgi-Open Pit process streams is not documented.	The operating experience with the Lurgi-Open Pit streams needs to be obtained.
		G,H	10,11	1	The technology is widely used in other industries. The estimated control efficiency of 99.7% appears reasonable.	The technology transferability needs to be verified.
Water and Foam Sprays (3.3-2, 3.3-10)	Particulates (fugitive)	I	1	2	The technologies are mentioned in the original DDP as the controls for fugitive dust, but the operating experience with the Lurgi-Open Pit process streams is not documented.	The operating experience with the Lurgi-Open Pit streams needs to be obtained.
		G,H	10,11	1	The technologies are widely used in other industries. The estimated control efficiencies of 85-98.5% for the foam sprays and 50% for the water sprays appear to be reasonable.	The technology transferability needs to be verified.
Retort Gas (3.3-3, 3.3-4, 3.3-5)	--	C	2	3	The retort gas composition was determined from a pilot plant experiment with the Tract C-a shale.	Scale-up data need to be obtained.
					Both H_2S and SO_2 have been reported to be present in the gas, but other sulfur species such as COS and mercaptans are not reported.	The presence or absence of COS, CS_2, mercaptans, etc., in the retort gas needs to be verified.
					Nitrogen compounds, other than NH_3, are not reported.	The presence or absence of organic amines, cyanides, etc., needs to be verified.

(Continued)

TABLE 7.1-1 (cont.)

Streams and Control Technologies (Figure No.)	Pollutant Controlled	Information Status[a]	Information Sources[b]	Reliability[c]	Remarks	Research Needs
Stretford (3.3-8)	H_2S	G,H,I	10,11	3	In this manual, the technology is used to treat the acid gases obtained during the retort gas purification.	
					The operating experience with the Lurgi retort gas is not documented.	The operating data from actual source testing need to be obtained.
					The technology has been tested recently with the retort gas from a pilot Modified In Situ retorting experiment, but the data are not yet available.	The pilot plant data need to be obtained and the transferability of the information to the Lurgi acid gases needs to be verified. Scale-up data may also need to be obtained.
					The technology is used commercially in other industries at a scale necessary to treat the Lurgi acid gases.	The technology transferability needs to be verified.
					Non-H_2S sulfur compounds may not be recovered efficiently with the technology.	The control efficiencies for COS, CS_2, mercaptans, etc., need to be determined.
					Excessive amounts of heavy organics tend to deteriorate the reagents and the quality of the sulfur product.	The impact on the efficiency of H_2S removal due to the presence of condensable organics in the feed needs to be quantified.
					Non-NH_3 nitrogen compounds may also degrade the reagents.	The impact of organic amines, HCN, etc., on the Stretford chemicals needs to be quantified.
					Excessive amounts of CO_2 in the feed may have an adverse affect on the H_2S removal efficiency.	The impact on the efficiency of H_2S removal due to excessive amounts of CO_2 needs to be quantified.
					According to the vendor information, an H_2S removal efficiency of 30 ppmv in the treated gas is achievable with a single absorber.	
Lurgi Flue Gas (3.3-3)	--	C,I	2	3	The flue gas data have been obtained from a pilot-scale experiment with the Tract C-a oil shale.	Scale-up data need to be obtained.

(Continued)

TABLE 7.1-1 (cont.)

Streams and Control Technologies (Figure No.)	Pollutant Controlled	Information Status[a]	Information Sources[b]	Reliability[c]	Remarks	Research Needs
					The SO_2 content of the flue gas is reported to be 30 ppmv. This amount appears to be too low based on the material and elemental balances. Adsorption of the SO_2 on the processed shale to form calcium and magnesium sulfates is given as the explanation for the low SO_2 emission.	The efficiency of SO_2 adsorption on the processed shale needs to be determined.
					The NOx content of the flue gas is reported to be 300 ppmv. Based on the material and elemental balances, this amount appears to be too low.	The actual NOx content of the flue gas needs to be determined.
					Only 10% of the fuel-based nitrogen in the processed shale is reported to be converted to NOx, while 90% is converted to elemental nitrogen. Approximately 50% of the fuel-based nitrogen is normally converted to NOx.	The conversion of the fuel-based nitrogen to NOx needs to be quantified. Also, the extent of thermal fixation of the atmospheric nitrogen needs to be determined.
					Data on trace elements and several criteria pollutants are not documented.	The data on trace elements and criteria pollutants need to be obtained from actual source testing.
Electrostatic Precipitator (3.3-3)	Particulates	C,I	2	3	An electrostatic precipitator to remove the particulates from the flue gas has been suggested in the modified DDP for Tract C-a.	
					The operating experience with the Lurgi pilot plant has been obtained.	Scale-up data need to be obtained.
		G,H	10,11	2	The technology is used commercially in the utility industry at a scale necessary to treat the Lurgi flue gas.	The technology transferability needs to be verified.
					The particulate removal efficiency depends upon the resistivity of the processed shale and the temperature of the flue gas stream.	The effect of variations in the shale grade on the resistivity of the particulates needs to be quantified.
					Moisture in the flue gas generally decreases the resistivity, thus increases the control efficiency.	The relationship between the moisture content of the flue gas and control efficiency needs to be studied.

(Continued)

TABLE 7.1-1 (cont.)

Streams and Control Technologies (Figure No.)	Pollutant Controlled	Information Status[a]	Information Sources[b]	Reliability[c]	Remarks	Research Needs
Fiberglass Fabric Baghouse	Particulates	G,H,I	10,11	2	The fiberglass baghouses have a much higher temperature limit than conventional baghouses, but the operating experience with the Lurgi flue gas is not documented.	The feasibility and efficiency of the technology for the flue gas need to be determined. Also, the effect of temperature needs to be studied.
					The technology is used in other industries. A particulate control efficiency comparable to that obtainable with conventional baghouses appears to be achievable.	The technology transferability needs to be verified.
Fugitive Hydrocarbons	--	I	10	2	The fugitive hydrocarbons are estimated from the properties of the oil products.	
Floating Roof Tanks	Hydrocarbons	I	10	2	Double-sealed, floating roof storage tanks have been provided for volatile product storage.	
		G,H	10,11	1	Floating roof storage tanks are used commercially for oil storage.	
Maintenance	Hydrocarbons	G,H	10,11	1	Routine maintenance of valves, pumps, etc., is a commonly used operational practice to control the hydrocarbon leakage.	
Catalytic Converters	Hydrocarbons, CO	G,H	10,11	1	All diesel-powered machinery is equipped with catalytic converters to control hydrocarbon and CO emissions. The catalytic converters are a commonly used technology.	
Gas Liquor (3.3-4)	--	C,I	1	3	The composition of the gas liquor has been determined from the pilot experiment with the Tract C-a shale.	Scale-up data need to be obtained.
Oil/Water Separator (3.3-4)	Oils and Greases	G,H,I	10,11	2	The operating experience with the gas liquor is not documented.	The feasibility and efficiency of the technology for removal of oils and greases from the gas liquor need to be evaluated.
					The technology is used commercially in other industries.	The technology transferability needs to be verified.

(Continued)

TABLE 7.1-1 (cont.)

Streams and Control Technologies (Figure No.)	Pollutant Controlled	Information Status[a]	Information Sources[b]	Reliability[c]	Remarks	Research Needs
					Oil emulsions may not be controlled by the separator. Addition of chemicals or heating the water may be necessary to break the emulsion.	The potential of forming oil emulsion in the gas liquor needs to be evaluated.
Ammonia Recovery Unit (3.3-9)	NH_3	G,H,I	4,10	2	The operating experience with the oil shale process waters is not documented.	The feasibility and efficiency of the technology for the Lurgi gas liquor need to be evaluated.
					The technology is used commercially in other industries.	The technology transferability needs to be evaluated.
					Dissolved organics in the gas condensate may have a detrimental impact on the efficiency of ammonia recovery and the quality of the product.	Dissolved organics in the gas condensate and their impact on the efficiency of the technology need to be estimated.
Carbon Adsorption (CA) (5.2-17, 5.2-25)	Dissolved Organics	G,H,I	5,10,11	3	The technology is used commercially in the treatment of industrial and municipal wastewaters. The operating experience with oil shale effluents is not documented. In this manual, the technology is used for polishing the stripped gas liquor before it can be used in the cooling tower. A 50% reduction in the organics appears to be achievable with this technology.	The feasibility and efficiency of the CA treatment for the Lurgi gas liquor need to be evaluated and/or the technology transferability needs to be verified.
Cooling Tower (3.3-11, 5.2-17)	Dissolved Solids	G,H,I	6,10,11	2	The cooling tower is a commonly used technology. It can be used to control the dissolved solids in the process waters if the volatile components have been removed previously and the water quality is suitable as the makeup to the cooling tower. In this manual, first the volatile components in the gas liquor are removed by steam stripping in the ammonia recovery system, then the organics are removed by adsorption on carbon. The water thus treated is evaporated in the cooling tower and the dissolved solids are concentrated in the cooling tower blowdown.	The feasibility and efficiency of the cooling tower for the stripped gas liquor need to be evaluated and/or the technology transferability needs to be verified.

(Continued)

TABLE 7.1-1 (cont.)

Streams and Control Technologies (Figure No.)	Pollutant Controlled	Information Status[a]	Information Sources[b]	Reliability[c]	Remarks	Research Needs
Solar Evaporation Pond (5.2-17)	Dissolved Solids	G,H,I	10	2	The technology is commonly used for concentrating the wastewaters. Solar energy incident on an open evaporation pond is used to evaporate the water. The precipitated salts may be removed periodically. In this manual, the stripped gas liquor after the carbon adsorption and cooling tower treatments is concentrated further in the solar evaporation pond. Sufficient storage capacity and surface area are provided to hold the water without overflowing during the low-evaporation, high-precipitation months.	Characterization and disposal approaches for the precipitated salts need to be evaluated.
Mine Water (3.3-2)	--	C	3	2	The composition of the water from the upper and lower aquifers has been determined from the drilling and pumping tests on Tract C-a. Based on the storage coefficients and transmissivity data, it was estimated in this manual that 43% of the total mine water was contributed by the upper aquifer and 57% was contributed by the lower aquifer. The average mine water flow rate was estimated to be 16,500 gpm, although the flows from both aquifers are quite variable. The water quality also varies considerably within an aquifer and between the two aquifers.	Additional data on the aquifer water quality and flow rates may need to be obtained to assess potential reuse, treatment, and disposal options for the excess mine water.
Reverse Osmosis (RO) (5.2-11, 5.2-12)	Dissolved Organics and Inorganics	G,H,I	7,10	3	The operating experience with the mine water is not documented, but the technology is used commercially in other applications. In this manual, the technology is applied to the excess mine water for the removal of bulk dissolved solids. Approximately 90-99% of the dissolved inorganics can be removed by the technology. The removal efficiency for organics may be somewhat lower. The treated water is cleaned further so that it can be discharged and the rejected material is used for processed shale moisturizing.	The feasibility and efficiency of the technology for the mine water need to be evaluated and/or the technology transferability needs to be verified.

(Continued)

TABLE 7.1-1 (cont.)

Streams and Control Technologies (Figure No.)	Pollutant Controlled	Information Status^a	Information Sources^b	Reliability^c	Remarks	Research Needs
Boron Adsorption (5.2-11, 5.2-13)	Boron	I	10	4	This is an ion-exchange technique involving a resin which is specific for boron. The operating experience with mine water is not documented. In this manual, the technology is applied to the RO treated water to remove the boron in order to meet discharge criteria.	The feasibility and efficiency of the technology for the mine water need to be evaluated.
Phenol Adsorption (5.2-11, 5.2-13)	Phenol	I	10	4	This is also an ion-exchange technique involving a resin which is specific for phenol. The operating experience with the mine water is not documented. In this manual, the technology is applied to the excess mine water after it has been treated by the RO and boron adsorption technologies. The treated water is then discharged on the surface.	The feasibility and efficiency of the technology for the mine water need to be evaluated.
Aeration Pond (5.2-11)	Organics and Alkalinity	G,I	10	2	With this technique, the wastewater is aerated by passing air or pure oxygen through it. This process affords decomposition of the chemically oxidizable organic matter as well as provides the oxygen for the biological growth to carry out biooxidation. Some oxidizable salts of heavy metals can also be precipitated out. In this manual, the technology is applied to the RO treated excess mine water. The aerated water is discharged on the surface.	The feasibility and efficiency of the technology for the mine water need to be evaluated.
Reinjection System	--	G,I	8,10	2	The technology is used for the deep well injection of some oil brine wastes, but the operating experience with the excess mine water on Tract C-a is not documented. In this manual, the excess mine water is first clarified in an enclosed clarifier, then injected into the upper aquifer.	The feasibility and efficiency of the technology for the mine water at Tract C-a need to be evaluated and/or the technology transferability needs to be verified.

(Continued)

TABLE 7.1-1 (cont.)

Streams and Control Technologies (Figure No.)	Pollutant Controlled	Information Status^a	Information Sources^b	Reliability^c	Remarks	Research Needs
Solid Wastes (3.3-10)	--	C	2,10	3	The Lurgi processed shale composition has been derived from the pilot plant information on the Tract C-a shale and the material and elemental balances.	Scale-up data need to be obtained.
		B	9	3	Some physical properties of the Lurgi processed shale from Tract C-a have been measured in laboratory testing.	Scale-up data need to be obtained.
		B	9	3	The quality of the leachate from the Lurgi processed shale has been determined in a laboratory experiment.	Scale-up data need to be obtained.
		I	1	2	Large quantities of the overburden and subore are produced during mining. The physical and chemical characteristics of these solid wastes have not been determined. The wastes are disposed of along with the processed shale.	The physical and chemical properties of the overburden and subore need to be determined. If these wastes are to be mixed with the processed shale, then the impact on the properties of the processed shale should be evaluated.
		I	10	4	Cooling tower blowdown, boiler blowdown, boiler feedwater treatment regeneration waste, mine water clarifier sludge, storm runoff, service and fire water, etc., are combined to form the processed shale moisturizing water.	The extent to which the process wastewaters need to be treated before mixing with the processed shale needs to be determined. Changes in the physical and chemical properties of the solid wastes due to the mixing of various plant wastewaters also need to be determined.
Open Pit Backfilling (3.3-10)		A	1,10	4	Backfilling of the open pit with the solid wastes, after the pit has been developed to a sufficient size, is mentioned in the original DDP for Tract C-a, but the design details are not given.	The issues associated with pit configuration, fill slope, logistics of simultaneous mining and backfilling, etc., need to be addressed by a detailed engineering analysis specifically tailored for the development site. Careful procedures for waste disposal and project shutdown need to be developed, keeping in perspective the potential of resuming open pit mining in the future.

(Continued)

TABLE 7.1-1 (cont.)

Streams and Control Technologies (Figure No.)	Pollutant Controlled	Information Status^a	Information Sources^b	Reliability^c	Remarks	Research Needs
					Placement of the wastes in the path of the two intercepted aquifers may create the potential for groundwater contamination after the mine dewatering is stopped.	The groundwater contamination potential needs to be assessed. The effectiveness of liner materials to isolate the waste from the groundwater needs to be evaluated. The advantages and disadvantages of mixing the wastes versus keeping them segregated need to be evaluated from the operational as well as environmental viewpoint. Means of reestablishing the aquifers need to be investigated. Long-term impacts of combining the aquifers in the pit need to be evaluated on the basis of water quality, recharge rate, regional usage, etc.
Runoff Diversion Sumps and Pumps	Leachable Compounds	I	10	2	During the backfilling operation, the runoff from the waste pile and the pit walls is gathered in the collection sumps located at the junction of the fill and walls. It is then pumped to the surface for eventual use in processed shale moisturizing. After the project shutdown, the runoff is allowed to flow into the pit.	
Dust Control	Particulates	I	10	2	The control of fugitive dust generated during waste transport and placement is achieved by water and foam sprays and by paving the haul roads.	Alternate systems for dust control, such as application of chemical binders and asphaltic emulsions, need to be evaluated.

(Continued)

TABLE 7.1-1 (cont.)

Streams and Control Technologies (Figure No.)	Pollutant Controlled	Information Status[a]	Information Sources[b]	Reliability[c]	Remarks	Research Needs
Reclamation and Revegetation	Leachable Compounds, Particulates,etc.	I	10	2	Grubbing, stripping, and clearing of the area is performed as part of the mining activities. The completed surface of the landfill is covered with soil and vegetated. The operating experience with revegetating the Lurgi processed shale is not documented.	Reestablishment of the vegetation on the landfill needs to be studied on a long-term basis.

[a] Information Status:

A Conceptual analysis.
B Laboratory, bench-scale studies--oil shale or similar industry.
C Pilot plant studies--oil shale or similar industry.
D Semi-works studies--oil shale or similar industry.
E Commercial-scale studies--oil shale or similar industry.
F Pilot-scale studies--related industries.
G Commercial-scale studies--related industries.
H Vendor provided information.
I Engineering calculations.

[b] Information Sources (detailed source information can be found in Section 8, References):

1 Gulf Oil Corp. and Standard Oil Co. (Indiana), March 1976.
2 Rio Blanco Oil Shale Co., February 1981.
3 Gulf Oil Corp. and Standard Oil Co. (Indiana), May 1977.
4 U.S.S. Engineers and Consultants, Inc., April 1978.
5 Cheremisinoff and Ellerbusch, 1978.
6 Hart, June 11, 1973.
7 Hicks and Liang, January 1981.
8 Mercer, Campbell and Wakayima, May 1979.
9 Woodward-Clyde Consultants, October 13, 1980.
10 Engineering calculations (DRI, SWEC, WPA).
11 Vendor estimates.

[c] Reliability:

1 Information is judged to be applicable, no problems envisioned.
2 Information applicable, but some design or scale-up problems may be encountered.
3 Information applicable, but significant design or scale-up problems may be encountered.
4 Information may be applicable, but both design as well as scale-up problems may be encountered.
5 Information may not be applicable without major design and scale-up modifications.

Source: DRI based on the references listed in footnote b.

SECTION 8

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